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PB-288 162

Operational Results for the Piscataway Model 5 MGD AWT Plant

Washington Suburban Sanitary Commission, Hyattsville, MD

Prepared for

Municipal Environmental Research Lab, Cincinnati, OH

Sep 78

United States
Environmental Protection
Agency

Municipal Environmental Research
Laboratory
Cincinnati OH 45268

EPA-600/2-78-172
September 1978

Research and Development

PB 288162

EPA

Operational Results for the Piscataway Model 5 MGD AWT Plant

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TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-600/2-78-172	2.
3. RECIPIENT'S ACCESSION NO. PB288162	
4. TITLE AND SUBTITLE: OPERATIONAL RESULTS FOR THE PISCATAWAY MODEL 5 MGD AWT PLANT	5. REPORT DATE: September 1978 (Issuing Date)
	6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S): Thomas P. O'Farrell, Robert A. Menke	8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS: Washington Suburban Sanitary Commission, Hyattsville, Maryland 20781	10. PROGRAM ELEMENT NO. BC611
	11. CONTRACT/GRANT NO. S802943
12. SPONSORING AGENCY NAME AND ADDRESS: Municipal Environmental Research Laboratory-Cinti,OH, Office of Research and Development, U.S. Environmental Protection Agency, Cincinnati, Ohio 45268	13. TYPE OF REPORT AND PERIOD COVERED: Final (4/72 - 6/74)
	14. SPONSORING AGENCY CODE: EPA/600/14

15. SUPPLEMENTARY NOTES

Project Officer: Sidney A. Hannah, (513/684-7651)

16. ABSTRACT

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17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Sewage Treatment*, Chemical Removal (Sewage Treatment)*, Activated Carbon Treatment, Coagulation, Clarification, Filtration, Water Pollution	Physical-Chemical Treatment, Tertiary Treatment	13B

18. DISTRIBUTION STATEMENT: RELEASE TO PUBLIC	19. SECURITY CLASS (This Report): UNCLASSIFIED	21. NO. OF PAGES: 108
	20. SECURITY CLASS (This page): UNCLASSIFIED	22. PRICE: PC A06/MF A01

EPA Form 2220-1 (Rev. 4-77)

* U.S. GOVERNMENT PRINTING OFFICE: 1978— 657-060/1477

EPA-600/2-78-172
September 1978

OPERATIONAL RESULTS FOR THE PISCATAWAY MODEL 5 MGD AWT PLANT

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the prevention, treatment, and management of wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

Conventional treatment of municipal wastewater produces an effluent that may need additional treatment if a high quality effluent is required for discharge or reuse. A number of tertiary treatment processes have been developed and evaluated at laboratory and small pilot scale but at the inception of this project, had not been adequately demonstrated at full scale. This publication reports the performance of a 5 mgd tertiary treatment plant using lime clarification, dual media filtration and granular activated carbon adsorption.

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ABSTRACT

A 5 mgd tertiary wastewater treatment plant was constructed to demonstrate treatment of effluent from a 5 mgd step aeration activated sludge plant. The two-stage high lime process with intermediate recarbonation, filtration and activated carbon adsorption operated at the design rate for 36 days between two failures of the reactor clarifiers. A single-stage low lime process with filtration and activated carbon adsorption operated for 89 days. The combined secondary and tertiary treatment removed > 97% of BOD, TSS and P in the raw wastewater. Capital cost of the 5 mgd two-stage high lime system was 4.7 million dollars and operating costs were estimated as 36 cents per 1000 gallons of wastewater.

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CONVERSION FACTORS

The following factors convert the English units used in this report to the SI metric unit in popular usage in water engineering practice.

English Unit	X	Multiplier	=	Metric Unit (SI)
Acre		4,047		m^2
Btu		1.055		kJ
cu ft		0.028		m^3
°F		0.555(°F-32)		°C
ft		0.3048		m
gal		3.785		ℓ
gpm		0.0631		ℓ/sec
gpm/sq ft		40.7		ℓ/min m^2
hp		0.7454		kW
in		2.54		cm
lb		0.454		kg
mil gal		3,785		m^3
ton		907		kg

ACKNOWLEDGEMENTS

We express our sincere appreciation to Herman Keys, Senior Plant Operator, and his staff of Commission Plant Operators for their dedication and initiative during start up and testing periods.

We thank the Commission's Herb Johnson, Jr., and other Electrical Maintenance Personnel for their start up assistance and maintenance service.

Also appreciation to Raul Celerio, Chief Chemist, and his laboratory personnel for testing and analysis of process samples and special gas tests.

We thank Lam Lim and Dave Thorne, Supervisors of the Secondary Plant for their continuous involvement with the plant and the flow coming to the Model Plant.

EPA's Blue Plains Pilot Plant should also be mentioned for sample analysis and technique support during the testing period.

I. INTRODUCTION

The WSSC Wastewater Treatment Facility is located in Piscataway, Maryland, south of Washington D. C., at the inlet to Piscataway Bay. The effluent from the plant discharges into the marshlands of Piscataway Bay and as a result, high quality effluent is required. Ultimately the effluent from the facility will flow via an underground pipeline across federally owned property and discharge into the deep water of the Potomac River. To date, construction of the pipeline has not been completed and the WSSC has made a maximum effort to reduce the load to Piscataway Bay. The secondary treatment facilities at Piscataway have undergone three major construction phases which include:

1. Construction of a 5 mgd step aeration activated sludge system in 1967.
2. Completion of 28.6 acres of polishing ponds in 1970.
3. Expansion of the step aeration system to 30 mgd and the installation of fluid solids incinerators in 1974.

The plant was first placed in service in November 1967.

Early in 1966, the Environmental Protection Agency, formerly the Federal Water Pollution Control Administration, and the Washington Suburban Sanitary Commission entered into a joint agreement for the construction and operation of a tertiary treatment facility. The "Joint FWPCA-WSSC Model Advanced Waste Treatment Plant" was to be located in Piscataway, Maryland, at the 313 acre site of the WSSC secondary facilities. The purpose of the plant was to demonstrate by advanced waste treatment techniques the removal of COD, BOD, suspended solids, phosphorus and refractory organics from secondary municipal wastewaters.

A 5 mgd step aeration activated sludge plant was under construction and was scheduled to be completed in late 1967 at the Piscataway site. The plant was to treat a mainly domestic wastewater from the rapidly expanding Prince Georges County, Maryland. The wastewater is pumped from sanitary sewers located in the Broad Creek, Swan Creek and Piscataway Creek drainage areas in Southern Prince Georges County.

In addition to the available plant site, the decision to construct the plant at Piscataway was also based on the increased interest in reducing the pollution load to the Potomac River Estuary. The Research and Development Program of the FWPCA had demonstrated in the laboratory and at small pilot-plant scale the technical feasibility of improving carbon and phosphorus removals from municipal waste discharges. In 1967, the only tertiary treatment system in operation was the 2.5 mgd facility at Lake Tahoe, California.

At that plant, alum was being used for chemical clarification. It was decided that full-scale operation was necessary to determine process reliability and to secure accurate cost information for the construction and operation of a tertiary treatment plant.

In December, 1966, a grant (WPRD 62-01-67) was awarded to the Washington Suburban Sanitary Commission by the FWPCA for the design and construction of a 5 mgd full-scale tertiary treatment plant. The plant was conceived to include: chemical clarification with lime, recarbonation, dual-media filtration, activated carbon adsorption, lime recovery, and activated carbon regeneration. The total estimated project cost was 2.2 million dollars with a Federal share of 1.65 million dollars.

Affirmative action on the project was delayed until July 1968, when a design entitled "Process Design for the Model Advanced Waste Treatment Plant-Piscataway, Maryland," was forwarded to WSSC by the FWPCA. The design included the two-stage high lime process with intermediate recarbonation and lime recovery by thermal recalcination, filtration, and activated carbon adsorption with thermal regeneration. Updated information, which had been obtained at the EPA-DC Pilot Plant in Washington, D.C., treating similar low alkalinity wastewater, was included in the design. At that time it was anticipated that the previously estimated 2.2 million dollars would be insufficient to cover the costs of a 5 mgd Advanced Waste Treatment Plant. Shortly after the process design was submitted to the WSSC, a design engineer was selected by WSSC with instructions to prepare a preliminary construction design, based on the submitted R&D process design, and a cost estimate. In January 1969, the engineer estimated the cost of the project at 3.2 million dollars. In June 1969, a supplementary grant (17080 DZY) was awarded to WSSC for $750,000, thus increasing the Federal share to 2.4 million dollars. Because of the limited available funds, the engineer was instructed to reduce costs where possible. The area most affected by the cost reduction was the elimination of duplicate equipment that would be necessary to ensure continuous operation.

The size of the plant was selected as 5 mgd. The final construction drawings and specifications were completed in May 1970. Review of the bid responses in September 1970, showed that the lowest bid was approximately 4.5 million dollars. An agreement was reached with the EPA Region III Construction Grants Division whereby Federal costs in excess of 2.4 million dollars would be paid by the Construction Grants Division as part of Project WPC-Md-233. Contracts for the construction of the 5 mgd tertiary facility were awarded in November 1970, with an expected completion date of January 1972. The final construction and initial operation of the plant were, however, delayed by many factors.

In April 1972, an Environmental Protection Agency grant was awarded to the WSSC for one year of operation of the tertiary facility. It was anticipated that the system would operate from July 1972, through June 1973. However, because of construction delays, start-up of the operation was delayed until January 1973. Mechanical failures in the system initially prevented continuous operation, thus allowing the operating grant to be extended until June 1974, without additional funds.

II. CONCLUSIONS

This publication reports results from the operation of a 5 mgd tertiary wastewater treatment plant used to upgrade the quality of effluent from a conventional secondary plant using step aeration activated sludge. The tertiary plant used either two-stage high lime or single-stage low lime followed by dual-media filtration and granular activated carbon adsorption. Major conclusions from the demonstration project are as follows:

1. The high lime process using an average dosage of 257 mg/l CaO and 18 mg/l $FeCl_3$ significantly reduced residuals of BOD, TSS and P in the secondary effluent. BOD was reduced from 16.5 mg/l to 4.0 mg/l; TSS was reduced from 27.5 mg/l to 2.5 mg/l and P was reduced from 3.50 mg/l to 0.10 mg/l.

2. The low lime process using an average dosage of 113 mg/l CaO and 25 mg/l $FeCl_3$ produced removals of BOD, TSS and P comparable to those obtained with the high lime process.

3. Tertiary treatment did not significantly affect total N residuals in the secondary effluent.

4. Carbon losses from regeneration of three columns under less than optimum conditions were estimated as 8-10%.

5. Operating costs for tertiary treatment in this demonstration plant were in the range of 29-36 cents per 1000 gallons. These are considered to be unusually high because data are based on the startup period when the plant was not at optimum efficiency and operators were not familiar with the plant.

6. A highly competent staff is required to successfully operate a complex tertiary wastewater treatment plant. Extra efforts should be made to select, train and retain personnel.

III. DETAILED DESCRIPTION OF SECONDARY TREATMENT FACILITY

The primary and secondary treatment facilities at Piscataway consist of two parallel systems (capacity = 5 and 25 mgd) with a common solids handling system. Raw wastewater is presently pumped to the plant from four pumping stations with the total capacity of 75 mgd. Although the total feed system has the capacity to feed 75 mgd, neither the sewage taps nor the plant has the capacity to collect or treat this volume of wastewater.

The flow enters the plant at the distribution structure via two force mains. The recycle of overflows and filtrate from the solids handling system also enters the distribution structure. From the distribution structure, the flow is split for delivery to three aerated grit chambers. A schematic diagram of the liquid treatment facilities is shown in Figure 1. Although the entire secondary plant is presented, only those unit processes associated with the 5 mgd system feeding the Model Plant will be discussed. The 5 mgd secondary system as operated during the grant is shown in Figure 2.

The raw sewage is manually split and fed to the three grit chambers. The effluent from the single grit chamber for the 5 mgd system passes through two 3/8" barminutors equipped with automatic rotating cutters. Four centrifugal pumps, each rated at 1750 gpm, are piped to the open channel following the barminutors to provide a constant flow of 5 mgd through the 5 mgd secondary section into the Model Plant. Since the three chambers are interconnected, all flow in excess of 5 mgd is diverted to the two grit chambers in the 25 mgd system.

The degritted effluent is split at the primary inlet well and fed to two parallel primary settling tanks. Each tank with an outside diameter of 60 ft is equipped with both bottom scrapers and surface skimmers. At a flow of 2.5 mgd per tank, the loading to each tank is 890 gal/day/sq ft with a detention time of 1.7 hours. The primary solids are wasted by gravity to the primary solids collection well. At the collection well, the primary solids are combined with the scum from the surface skimmers and flow by gravity to the sludge thickener. Wasting of primary solids is a manual operation without automatic flow measurement.

Primary effluents from the two clarifiers are combined and flow by gravity to the step aeration basins. Two parallel reactors are provided with a common feed channel located in the center of the two reactors as shown in Figure 2. Primary effluent is fed to the quarter points of the reactors at a rate of Q/6 through each gate. Settled solids from the secondary clarifiers are recycled to the head of the two reactors to maintain an average MLSS concentration of 2000 mg/1. Normally, the recycle flow was maintained

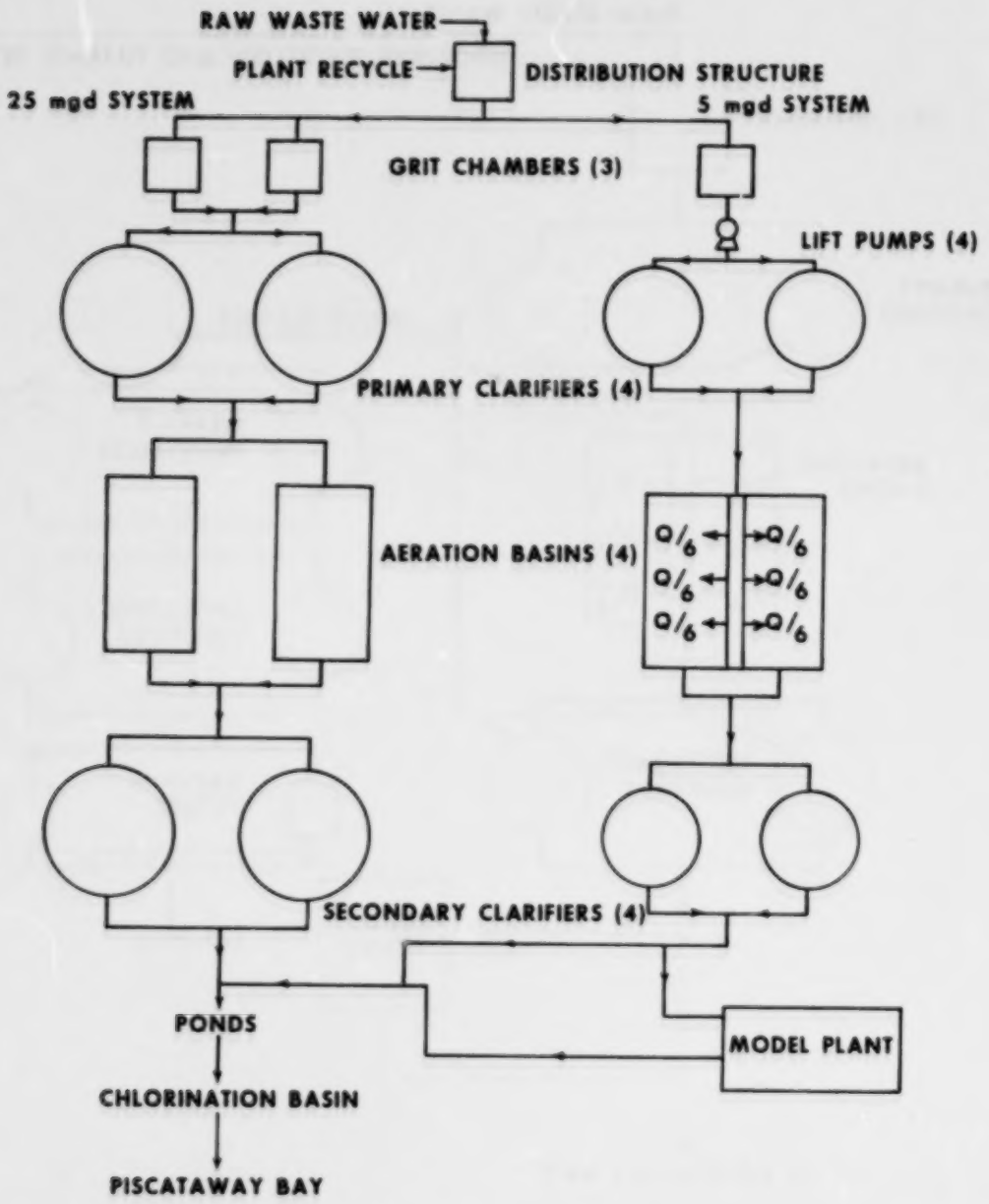

Figure 1. Flow schematic of the Piscataway secondary plant.

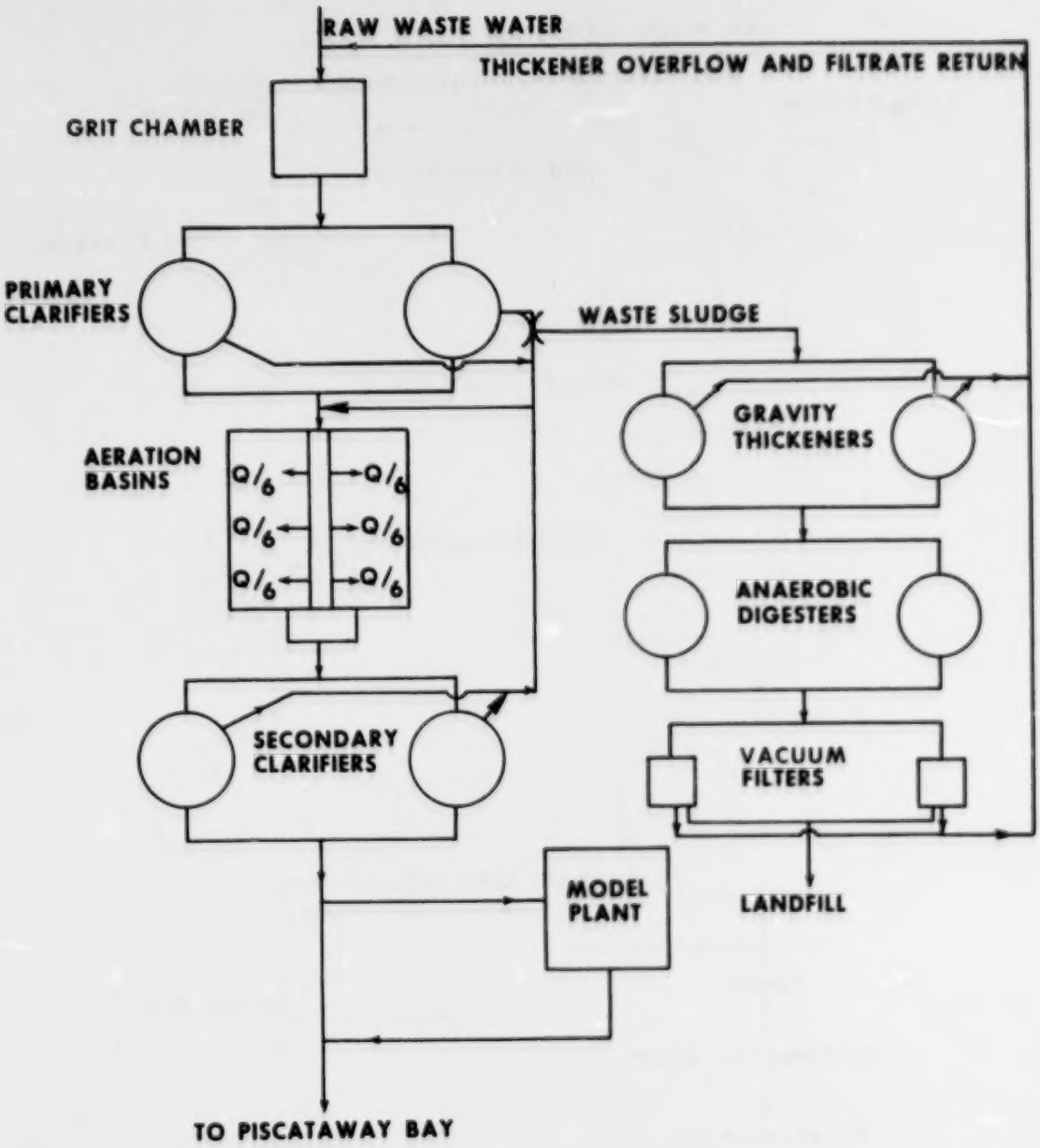

Figure 2. Schematic of the 5 mgd system of the Piscataway plant secondary plant.

at 33% of total flow. At a flow of 5 mgd, the detention time in the reactors is 4.5 hours. Air is supplied by diffusers located on the outside walls of the reactors. The dissolved oxygen concentration in the reactors is controlled by manually operated valves located above the water level. Normal operation called for control of the DO at approximately 2.5 mg/l O_2. The diffusers not only provide the required DO but also create a circular motion to promote mixing. Surface sprays are located in each reactor along the center feed channel.

The mixed liquor flows by gravity to the secondary settler inlet structure where the flow is split and fed to two settlers. At a flow of 2.5 mgd/settler, the units provide 2.75 hours of detention time at a surface loading of 650 gpd/sq ft. Centrifugal pumps return the settled solids to the reactors. Wasting of solids from the system is accomplished by diverting a portion of the recycled flow to the gravity thickeners. The effluents from the two clarifiers are combined and measured in a Parshall flume. The total flow or a portion of the total flow is sent to the Model Plant's inlet structure. Excess flow not reporting to the Model Plant is diverted through a Parshall flume to the polishing ponds which were installed at the plant as an interim upgrading step pending completion of the 25 mgd system.

The waste solids from both the primary and secondary settlers flow by gravity to the thickeners' inlet structure from which the flow is directed to four gravity thickeners; two basins 35 ft in diameter and two basins 55 ft in diameter. The gravity thickening system was designed for a total plant flow of 30 mgd. Since the total flow was less than 15 mgd during the test period, only two or three of the thickeners were in operation.

The underflow from the thickeners was pumped to two anaerobic digesters. The digesters were designed on a basis of 5 mgd wastewater flow but had to be used at higher loadings because of a Prince Georges County Council ban on the use of three fluidized bed solids incinerators. The WSSC was not allowed to operate the incinerators because of possible air pollution. The capacity of the digesters was the controlling factor in the solids handling system. Because of the limited capacity of the digesters, thickener overflow, high in suspended solids, was recycled to the inlet of the plant.

Following anaerobic digestion the underflow is vacuum filtered. The cake is trucked to a farm for spreading. Because of health requirements, only stabilized sludge can be vacuum filtered and disposed of on the land.

IV. TERTIARY TREATMENT

Operation of the tertiary treatment plant consisted of treating the secondary effluent from the 5 mgd step aeration system by lime clarification, dual-media filtration and granular activated carbon adsorption. The solids from the chemical clarification system were gravity thickened and dewatered by solid bowl centrifuges. The dewatered cake was recalcined in a multiple hearth furnace. The exhausted granular activated carbon was thermally regenerated in another multiple hearth furnace.

Lime Treatment

In lime clarification of relatively low alkalinity wastewaters, two options are available: two-stage high lime with intermediate recarbonation or single-stage low lime. Since the alkalinity of the Piscataway wastewater is less than 150 mg/l as $CaCO_3$ and adequate flexibility was incorporated into the design of the plant, both the single- and two-stage systems were evaluated.

A lime slurry, when added to a secondary wastewater, raises the pH of the liquid to produce chemical precipitation. Above pH 8.3, bicarbonate ions are converted to carbonate ions which react with the available calcium ions to precipitate calcium carbonate.

$$Ca^{++} + HCO_3^- + OH^- \longrightarrow CaCO_3 + H_2O$$

Above neutral pH, the calcium ions react with phosphate ions in the presence of hydroxyl ions to precipitate hydroxyapatite.

$$5Ca^{++} + 3H_2PO_4^- + 7OH^- \longrightarrow Ca_5(OH)(PO_4)_3 + 6H_2O$$

Millipore filtered samples from laboratory jar tests have shown that at pH 10 nearly all of the phosphate is precipitated, however, because of the lack of a coagulant aid, the precipitate is not readily removed by simple settling. By adding sufficient lime to increase the pH above 11.3, the magnesium ions, which are naturally present at between 3 and 6 mg/l in Piscataway wastewater, react to precipitate magnesium hydroxide, an excellent coagulant aid.

$$Mg^{++} + 2OH^- \longrightarrow Mg(OH)_2$$

With the aid of the magnesium hydroxide, the precipitated phosphorus and calcium carbonate coagulate and settle.

When the pH of a low alkalinity wastewater has been increased to above 11.3 by the addition of calcium hydroxide, an excess of calcium ions exists. The calcium ion concentration in Piscataway wastewater is increased from approximately 30 mg/l at neutral pH to approximately 100 mg/l at pH 11.5. The excess is produced because more calcium hydroxide is required to elevate the pH to 11.5 than there are available phosphorus and bicarbonate ions to react with the added calcium ions.

In the two-stage high lime system shown in Figure 3, lime slurry, wastewater and recycled settled solids are mixed in the rapid mix zone of the first clarifier to reach pH 11.5. The settled solids are recycled to increase the rate of precipitation. Following rapid mix, flocculation and settling, the effluent from the first clarifier flows to the recarbonation basin where gaseous carbon dioxide is added for precipitation of the excess calcium ions at pH 10.

$$Ca^{++} + 2OH^{-} + CO_2 \longrightarrow CaCO_3 + H_2O$$

Reducing the pH of the effluent from pH 11.5 directly to pH 7 by the addition of carbon dioxide resulted in solution of the calcium carbonate with an increase in the hardness of the water and the loss of the potentially recoverable calcium ions. The effluent containing the precipitated calcium carbonate is then fed to a second clarifier. The calcium carbonate is difficult to settle at pH 10 and a coagulant aid is required. Ferric chloride, added at a dosage of 5 to 10 mg/l as Fe^{+++}, has been successfully used as a coagulant aid.

$$Fe^{+++} + 3OH^{-} \longrightarrow Fe(OH)_3$$

Following the settling of the calcium carbonate in the second clarifier, carbon dioxide is added to the effluent to reduce the pH to 8 prior to filtration. The neutralization of the effluent prevents calcium carbonate scaling of the filter media.

$$OH^{-} + CO_2 \longrightarrow HCO_3^{-}$$

$$CO_3^{=} + CO_2 + H_2O \longrightarrow 2HCO_3^{-}$$

In the single stage lime system, shown in Figure 4, the pH of the wastewater is increased to 10.5 by the addition of lime in the rapid mix zone of the clarifier. As stated above, much of the phosphorus is precipitated along with the available carbonate. However, without the precipitation of magnesium hydroxide, good clarification does not occur and a coagulant aid is required. Ferric chloride is added for this purpose to the rapid mix zone of the single clarifier. The concentration of excess calcium ions in the effluent from the single-stage system is approximately 50 to 60 mg/l as Ca^{++} and recovery by the addition of carbon dioxide is not economically feasible. Following settling in the single clarifier, the effluent is neutralized to pH 8 and filtered. The single-stage lime system is not as effective as the two-stage for removing phosphorus.

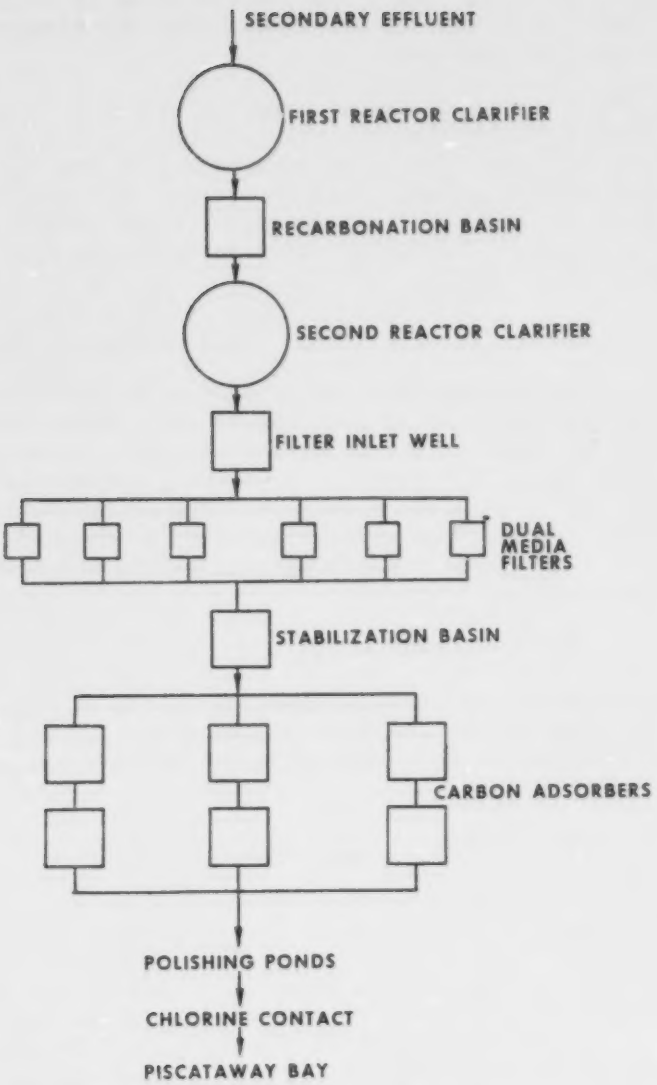

Figure 3. Two-stage high lime tertiary process.

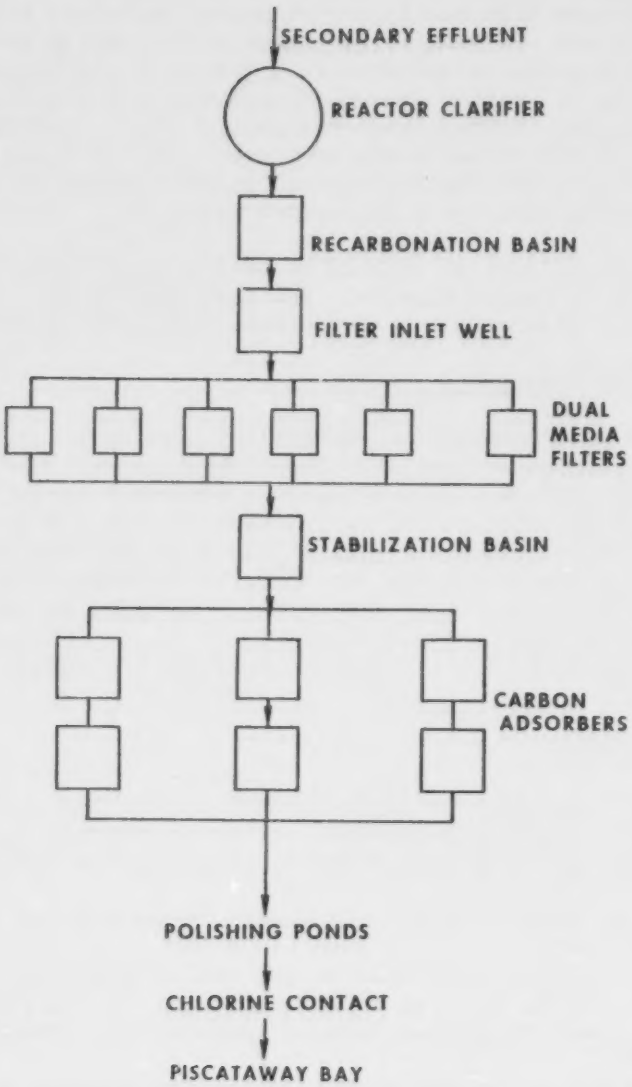

Figure 4. Single stage low lime tertiary process.

Neutralized effluent is applied to dual media gravity filters for the removal of particulate materials including phosphorus and organics. The filter media consists of 18 inches of anthracite coal (effective size 0.85 to 0.95 mm) over 6 inches of fine sand (effective size 0.40 to 0.45 mm). The 24-inch filter bed is supported by 12 inches of graded gravel (#10 mesh on top to 1 inch on the bottom). The coarse to fine gradation of the media produces in-depth filtration with the larger suspended particles being removed in the anthracite coal and the smaller particles being removed in the fine sand. The filters were normally operated at 3 gpm/sq ft and backwashed at a maximum of 20 gpm/sq ft, either on a predetermined time schedule or upon reaching terminal head loss. Surface wash was applied during backwash to cleanse the upper portion of the filter media to prevent the formation of mud balls. The backwash water for both the filters and carbon columns is stored and returned to the head of the plant at a controlled rate.

The filter effluent is reduced to pH 7.5 in the water stabilization tank by the addition of carbon dioxide. The reduction in pH is necessary to meet stream discharge standards and to optimize activated carbon adsorption.

Activated Carbon Adsorption

The final unit process in the tertiary treatment scheme is activated carbon adsorption. The effluent from the stabilization tank is pumped at a rate of 6.5 gpm/sq ft through a packed granular activated carbon bed which adsorbs soluble organic materials from the wastewater. The adsorbed organics serve as a food source for bacteria which multiply on the carbon to produce biological slimes. The biological activity, if controlled, can substantially increase the life of the activated carbon. In order to control the activity and to prevent excessive pressure losses through the packed carbon beds, backwash and surface wash of the carbon columns are necessary. The backwash rate is 15 gpm/sq ft.

Carbon Regeneration

Following exhaustion of the activated carbon, the carbon is removed from the column for thermal regeneration. It is transferred, at a controlled rate, into a multiple hearth furnace where the regeneration takes place in 4 stages:

1. The wet carbon is dried by simple evaporation at temperatures above 200°F.
2. Upon application of heat to the carbon grains at temperatures above 600°F, the high molecular weight impurities on the carbon will crack to produce gaseous hydrocarbons, hydrogen and water vapor.
3. The final regeneration step is the gasification of the residue from the pores of the carbon grains. This is accomplished using steam (approximately 1 lb steam/lb dry carbon) at temperatures between 1700 and 1850°F. The gaseous products of the reactions are carbon monoxide and hydrogen.

4. The regenerated carbon is finally cooled rapidly to ambient temperatures by water sprays.

Accurate control of the regeneration process is essential to maintain maximum gasification of the organic residue without causing thermal destruction of the granular carbon.

Lime Handling and Recovery

The chemical solids from the lime clarification system must be either recovered for reuse or subjected to ultimate disposal. Laboratory tests and material balance calculations show that at Piscataway in the single-stage system, the solids production is approximately 2 lb/1000 gal with a calcium carbonate concentration of 50%. The two-stage system will produce approximately 4 lb/1000 gal with a 75% calcium carbonate concentration. One of the objectives of this study was to determine the cost of solids handling and calcination both with and without recovery of the calcium carbonate.

For both land disposal or lime recovery, maximum dewatering of the sludge is required to reduce operating costs, i.e., trucking costs are based on total weight and heat requirements increase with the quantity of water to be evaporated in the furnace. In both the single stage and the two-stage systems, the sludge is pumped to a gravity thickener. The sludge from the second clarifier of the two-stage system is normally returned to the rapid mix zone of the first stage to increase the calcium carbonate concentration, thus improving settling. Wasting from the system occurs from the first clarifier. Solids are removed from the clarifier at 3-5% total solids and thickened to 10-15% by a gravity thickener. The overflow from the thickener is returned to the first clarifier and the underflow solids are pumped to a centrifuge for dewatering.

In dewatering the chemical sludges by centrifugation, two modes of operation are available: total solids recovery and two stage wet classification. The classification technique is used only when lime recalcination is employed. In the total solids recovery mode, organic polymers are injected into the centrifuge for maximum solids recovery. The centrate with a minimal solids concentration is returned to the head of the plant. The wet cake may be calcined for lime recovery or wasted. Thermal recalcination takes place in a multiple hearth furnace by increasing the temperature of the cake to 1850°F with an auxiliary fuel source to convert the calcium carbonate to calcium oxide.

$$CaCO_3 \longrightarrow CaO + CO_2$$

The gaseous by-product, carbon dioxide, is recovered and used in the recarbonation and stabilization tanks. The recalcined lime is slaked for reuse in the lime clarification system.

For the two stage wet classification mode, two centrifuges are operated in series with the first machine being operated for capture of the calcium carbonate in the centrifuge cake with the inerts reporting to the centrate. At higher feed rates and without the addition of organic polymers, approximately 95% of the calcium carbonate in the sludge reports to the cake. The centrate from the first centrifuge is pumped to the second machine where by the addition of organic polymer the inert solids are captured in the cake. The cake from the first machine is fed to the recalcination furnace and the cake from the second machine is trucked to a landfill with the final centrate being returned to the head of the plant. Since the cake from the first unit is high in calcium carbonate, the recalcined lime contains a high percentage of calcium oxide and wasting of the recalcined lime is not required. Additional lime, however, is required to maintain the high pH required in the lime clarification processes.

In the total recovery operation, a portion of the recalcined lime, equal to the inert build-up, is wasted from the system. Since calcium oxide is also wasted with the inerts, it is necessary to add lime which is equivalent to the amount wasted. In the two-stage system with secondary effluent, the make-up rate is approximately 25%.

V. DETAILED DESCRIPTION OF THE MODEL TERTIARY PLANT FACILITY

The Model Plant equipment is presented in Tables 1-3. The flow rate entering the plant is dependent on two operations. First, just past the aerated grit chambers of the secondary treatment plant are located four vertical constant speed centrifugal pumps each rated at 1,750 gpm at 5 ft total dynamic head. The lift pumps were installed to maintain a constant flow to the Model Plant. The second point of control is in the 30-inch influent pipe to the Model Plant where an automatic flow control valve was provided to control the influent flow rate to the Model Plant and isolate it from process flow in case of an extended power failure or Model Plant shutdown. A 14-inch magnetic flowmeter was provided to monitor the plant's flow and transmit a signal to the main control panel to be recorded and totalized. The flow signal is then transferred back to the automatic control valve to maintain a predetermined flow as set on the controller at the main panel. The inlet structure, two reactor clarifiers, recarbonation basin, filter inlet tank and gravity thickener are located outside while the remaining units are located inside the operations building.

Chemical Clarification

The flow of secondary effluent enters the influent structure which has the capability of directing the flow to either of the two reactor clarifiers. The flow pattern would depend on the preselected process to be evaluated as shown in Figures 3 and 4. The effluent from the inlet structure flows to the center draft tube of the first reactor clarifier, an 80 ft diameter x 16 ft side wall depth tank. In the draft tube, lime slurry, ferric chloride and/or polymer, settled solids and secondary effluent are combined under rapid mixing conditions. The settled solids are drawn off the bottom of the clarifier by the integral turbine mixer or may be recycled using an external variable speed centrifugal pump rated at 360 gpm.

The chemicals are applied by several methods. The lime handling system, as shown in Figure 5, is controlled by an automatic non-clog ball valve regulated by a signal sent to the control panel by a pH probe located in the flocculation zone. The feed valve automatically opens and closes at time intervals of one minute to prevent clogging. The 10% lime slurry is recirculated at 120 gal/min from the lime slurry storage tank to the clarifier through a series of 3 in steel lines and 2-1/2 in 100% dacron firehose and returned to the storage tank. Most of the hose was installed in an open trough to allow easy access for cleaning and replacement.

TABLE 1

DESIGN DATA FOR MODEL PLANT EQUIPMENT

REACTOR CLARIFIER #1

1 unit - 80 ft dia x 16 ft SWD
Total detention time: 2.74 hr
Surface overflow rate: 1300 gpd/sq ft @ 5.6 mgd

RECARBONATION BASIN - 2 units

15.2 ft x 15 ft x 19.5 ft
Volume: 4446 cu ft or 33,256 gal
Detention Time: 9.6 min @ 5 mgd each

REACTOR CLARIFIER #2

1 unit - 70 ft dia x 16 ft SWD
Total detention time: 2.17 hr
Surface overflow rate: 1630 gpd/sq ft @ 5.4 mgd

INLET WELL

21 ft x 15 ft x 12 ft SWD
Volume: 3780 cu ft or 28,274 gal
Detention time: 7.6 min

GRAVITY FILTERS - 6 units

Surface Area: 242 sq ft each
Filter rate: 3.1 gpm/sq ft
Backwash rate: 20 gpm/sq ft
Media

Top	Anthracite coal
	Sand
Bottom	Gravel

STABILIZATION BASIN - 1 unit

Volume: 34 ft x 17 ft x 12 ft SWD
6936 cu ft or 51,881 gal
Detention Time: 15 min

TABLE 1 (Cont'd.)

DESIGN DATA FOR MODEL PLANT EQUIPMENT

ADSORBER INFLUENT PUMPS

3 pumps - (centrifugal)
50 hp 2000 gpm each

ACTIVATED CARBON COLUMN ADSORBERS - 6 units

Volume: 15 ft dia x 29 ft SWD: 5125 cu ft or 38,335 gal
Detention time in bed: 18.3 min each
Backwash rate: 20 gpm/sq ft
16 ft of granular activated carbon on a 1 ft gravel bed

GRAVITY THICKENER - 1 unit

50 ft dia x 10 ft SWD
Total Detention Time: 18.7 hr
Surface overflow rate: 115 gal/day sq ft
Solids loading: 31 lb/day/sq ft

CENTRIFUGES - 2 units

Solid bowl super-D-canter
Size P-3400 30 hp Motor
Hydraulic capacity: 45 gpm
Scroll Speed: 3400 rpm

RECALCINATION FURNACE - 1 unit

Multiple hearth 6 levels
Design capacity total dry solids: 49,670 lb/day
Dia: 19 ft
Total solids by Weight: 20%-55%
Moisture content by weight: 80%-45%
Maximum temperature in any hearth: 2000°F
Operating Pressures: negative 0.1 to negative 0.3 inches of water column
Four middle hearths have 2 burners each
Fuel Supply: #2 fuel oil

TABLE 1 (Cont'd.)

DESIGN DATA FOR MODEL PLANT EQUIPMENT

CARBON REGENERATION FURNACE - 1 unit

Multiple hearth, 4 hearths plus afterburner section
Design Capacity: 4750 lb/day dry carbon
Maximum Furnace temperature on any hearth: 1860°F
Afterburner Temperature Maximum: 1500°F
Fuel Consumption: Maximum 4,000 BTU/lb product
Fuel Supply: Natural gas for four burners and
#2 Fuel oil for two burners in afterburner chamber

TABLE 2

MODEL PLANT PUMPS

PUMPS		
1. Clarification System		
2-Lime sludge recycling	centrifugal	360gpm
2-Lime sludge wasting	progressive cavity	150gpm
2. Filter and carbon adsorption system		
3-Adsorber pump	centrifugal	2000gpm
2-Backwash supplies	centrifugal	5000gpm
1-Filter decant (backwash effluent to headworks of Model Plant)	centrifugal	200gpm
1-Adsorber decant (backwash effluent to headworks of Model Plant)	centrifugal	200gpm
SOLIDS HANDLING		
1-Centrifuge feed	progressive cavity	55gpm
1-Furnace feed	progressive cavity	20gpm
2-Classification pump	progressive cavity	45gpm
CHEMICAL FEED		
4-Polymer	positive displacement	0.83gpm
4-Ferric Chloride	positive displacement	0.83gpm
1-Ferric Chloride (transfer 40% $FeCl_3$ to dilution tanks)	centrifugal	200gpm
2-Lime slurry transfer	centrifugal	40gpm
2-Lime slurry recirculating	centrifugal	120gpm
MISCELLANEOUS		
2-Utility supply water	centrifugal	350gpm
2-Operations building main sump pumps	centrifugal	350gpm

TABLE 3

MAJOR EQUIPMENT VENDORS

EQUIPMENT DESCRIPTION	EQUIPMENT VENDOR
Reactor-Clarifiers	Rex Chainbelt
Mixers	Mixing Equipment Co.
Dual Media Filters underdrains	F.B. Leopold Co.
Centrifuges	Sharples
Lime Recalcination Furnace	Nichols Corp.
Carbon Regeneration Furnace	Nichols Corp.
Lime Slaking System	Wallace & Tiernan Inc.

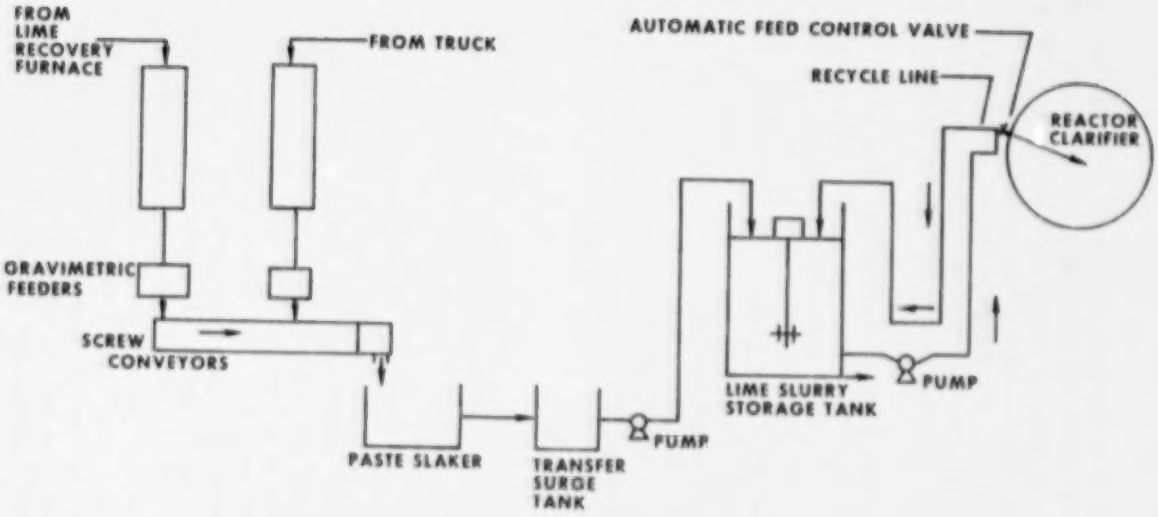

Figure 5. Lime handling system.

The pebble virgin lime is delivered to the plant by truck in bulk and stored in a 1240 cu ft storage tank. Both the pebble and recalcined lime systems are equipped with gravimetric feeders to add a constant weight of each material to the slaker. The lime is transported to the slaker via screw conveyors.

The paste type lime slaker, rated at 2000 lb/hr, converts the calcium oxide to calcium hydroxide by the addition of water. The chemical reaction is exothermic which increases rates of reaction. The paste is degritted and diluted prior to discharge to the slurry tank. The lime slurry is pumped from a 3 ft x 3 ft x 3 ft surge tank by a 200 gpm centrifugal pump to the 12,000 gallon lime slurry storage tank. The storage tank is equipped with a 15 hp mixer to maintain the solids in suspension.

In addition to the lime feed system, the plant is equipped with two 6,000 gallon ferric chloride storage tanks that receive 40% $FeCl_3$ solution via truck delivery. The 40% solution is pumped to one of the two 1500 gallon tanks where the solution is diluted to 10% prior to use. Four manually controlled diaphragm metering pumps, each rated at 50 gallons/hr, are provided to feed the ferric chloride. The polymer feed system consists of two 600 gallon solution tanks and four manually controlled diaphragm feed pumps each rated at 50 gallons/hr. The powdered polymer is received in 50 lb bags and fed to the tanks by an operator. The units are equipped with mechanical mixers. The dosages of these chemicals were varied as a function of the supernatant clarity and settleability of the suspended solids in the clarifier.

The reactor clarifiers, shown in Figure 6, consist of three major sections: draft tube, flocculation zone and clarification zone. The draft tube is located in the center section of the clarifier just below the turbine mixer and is used as a rapid mix zone. The process water, chemicals and recycled sludge are drawn by the mixer up and out into the flocculation zone. The impeller provides rapid mix to disperse the chemicals and to complete the chemical reactions and flocculation to promote particle growth. The detention time of the flocculation zone is estimated as approximately 20 minutes. Another function of the clarifier is to promote the settling of the precipitated solids. The process flow moves down inside the fiberglass cone separating the flocculation zone from the clarification zone. This baffle cone extends about half way down into the tank and the process flow moves through the flocculation zone and under the baffle. At this point the liquid and solids begin to separate as the solids continue their downward movement and the liquid begins to move upward. Testing of the baffle showed that its length was insufficient to dissipate the kinetic energy created by the mixer in the clarifier. The settled solids are collected by a plow rake mechanism and moved to a center hopper. Sludge wasting and recycle exits are located in the center hopper of the unit. To prevent scale accumulation the sludge was transported through glass-lined pipes from the hopper to the pumps. The clarified wastewater moves up towards the surface and into a series of orifices in the pheripheral effluent trough. The orifices vary in size in an attempt to achieve a uniform draw off from the clarifier. The total detention time in the first

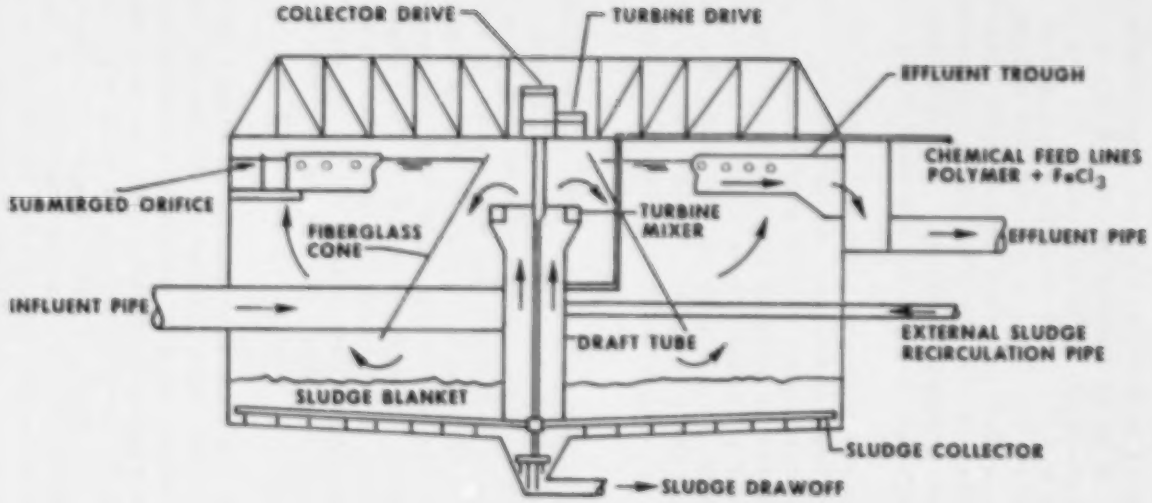

Figure 6. Cross section of reactor clarifier.

clarifier is 2.74 hr with an overflow rate of 1300 gpd/sq ft at a flow of 5.6 mgd.

From the first clarifier, the liquid flows through an open trough to the recarbonation basin for pH adjustment by the addition of carbon dioxide. The process flow enters the 32 ft x 15 ft x 12 ft tank and flows under and through a redwood baffle wall into the first of two carbon dioxide mixing zones. The first zone is used to accomplish approximately 75% of the required pH adjustment. A 30 hp turbine mixer and a 6-in flue gas feed line are provided in the first zone. The effluent from the first stage passes over a concrete baffle and into the equally sized second stage. The second stage, which is primarily for final adjustment of the pH, includes a 25 hp turbine mixer and a 4-in flue gas feed line. Both stages are equipped with submerged pH probes that control the flow of carbon dioxide to the separate stages by automatic gas regulating valves. The recorders and controllers for the pH control loops are located on the main control panel.

A 3-in sludge recycle line was located in the first stage recarbonation zone. During the two-stage operation, settled solids were recycled at a rate of 150 gpm from the center hopper of the second stage clarifier. The purpose of the recycled solids was to promote the precipitation of calcium carbonate and produce a more dense particle for settling. During the single-stage operation, calcium carbonate recovery was not required and the recarbonation basin was used to lower the pH below 8.3 where calcium carbonate is soluble. Sprays were installed to reduce foaming produced by the large volumes of flue gas introduced.

Gaseous carbon dioxide was provided from two sources; the recalcination furnace and a 25-ton liquid carbon dioxide storage tank. The liquid carbon dioxide source was used to supplement the carbon dioxide demand during periods of zero feed to the furnace and was placed in full service when the furnace was shut down. The concentration of carbon dioxide in the flue gas from the recalcination furnace varied from 5% at operating temperatures and zero feed to 12% at rated temperature and full feed capacity. At a 5% carbon dioxide concentration and wastewater flow of 5 mgd, additional carbon dioxide was required. The blending of two gases, at 5% CO_2 and 98% CO_2, respectively, produced pH control difficulties. The CO_2 feed system with automatic pH control had been designed to feed a 10% gaseous flow.

The effluent from the recarbonation basin flowed to the draft tube of the second stage clarifier. The purpose of the second stage clarifier was to capture the precipitated calcium carbonate produced during the two-stage operation. During the single-stage operation the unit was not used. The second clarifier is 70 ft in diameter with a side wall depth of 16 ft and an overflow rate of 1,630 gpd/sq ft at 5.4 mgd. The design of the internal mechanism and the chemical feed lines was identical to that for the first clarifier. During the two-stage operation, ferric chloride was pumped from solution tanks located in the operations building to the second-stage draft tube to improve chemical clarification.

Lime Recovery Operations

As stated earlier, the settled solids from the second stage clarifier were recycled to the recarbonation basin by a centrifugal pump. Excess solids from the second stage clarifier were wasted to the first stage clarifier via a variable speed progressive cavity pump. A similar pump was used to waste solids from the first stage clarifier to the gravity thickener. The thickener has an inside diameter of 50 ft and a side wall depth of 10 ft. Thickening of the chemical sludge is enhanced by a mechanical rake operated at 4.6 rph and powered by a 2 hp motor. The rake was designed to be raised or lowered as a function of the solids concentration. The overflow from the thickener flowed by gravity to the first stage clarifier. During normal operation, the overflow from the thickener was relatively free of solids. At the sludge production of the two-stage high lime system of 22,500 lb/day, the surface loading to the thickener was low at 11.5 lb/day/sq ft.

During the operation of the furnace or sludge wasting from the total system, the underflow from the gravity thickener was pumped to one of two solid bowl centrifuges for dewatering. The centrifuges were located in the operations building below the recalcination furnace. The centrifuges are powered by 30 hp motors and operate at a scroll speed of 3400 rpm.

The sludge is fed by an inner tube to approximately the center of the unit as shown in Fig. 7. As the sludge exits into the main drum area, the solids are moved slowly by a screw conveyor to the cake end of the machine. The liquid as it separates from the solids is mixed with polymer in order to capture additional solids. The polymer is added to the centrifuge by an additional inner tube just past the sludge addition point. The captured solids, because of their increased weight due to the polymer addition, move toward the cake end of the unit with the liquid moving toward the opposite end and discharging from the machine as the centrate. The centrate flows to the operations building's sump and is returned to the head of the model plant.

The arrangement of the two centrifuges allowed two different modes of operation of the solids dewatering system, total capture and two-stage wet classification. The operation of the centrifuge as explained above and shown in Fig. 8 provided total capture of the solids. The purpose is for maximum capture of the total solids entering the centrifuge and produces a centrate containing low solids. Classification for separation of the calcium carbonate from the inert solids is accomplished using two centrifuges in series as shown in Fig. 9. Thickener underflow is fed without polymer addition to the first centrifuge where most of the calcium carbonate reports to the cake and is fed to the furnace. The centrate from the first unit containing inerts and some calcium carbonate flows to a holding tank and is pumped to the second centrifuge where by the addition of polymer the solids are captured in the cake. The centrate from the second centrifuge returns to the head of the plant via the operations building's sump pumps.

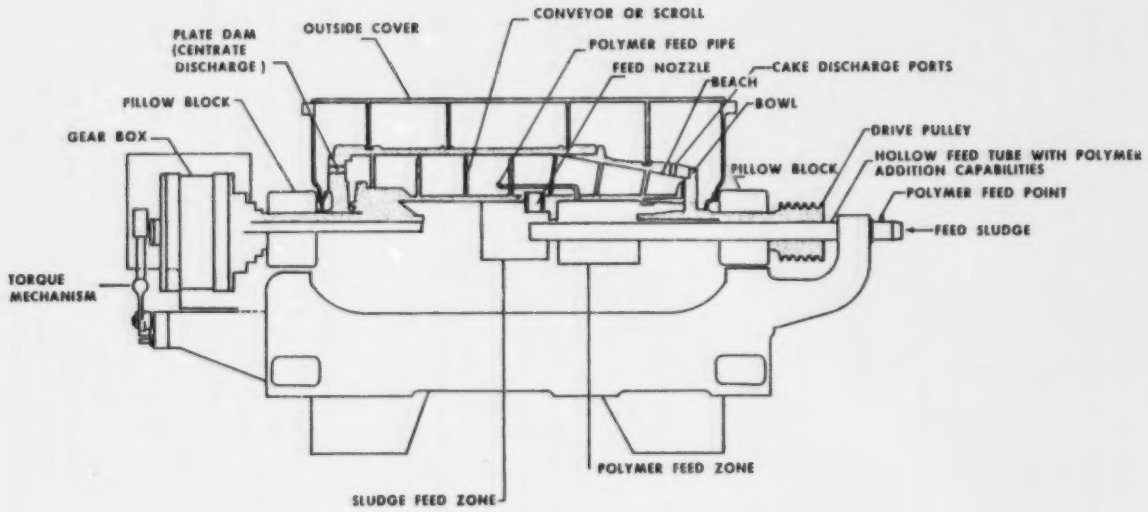

Figure 7. Solid bowl centrifuge section.

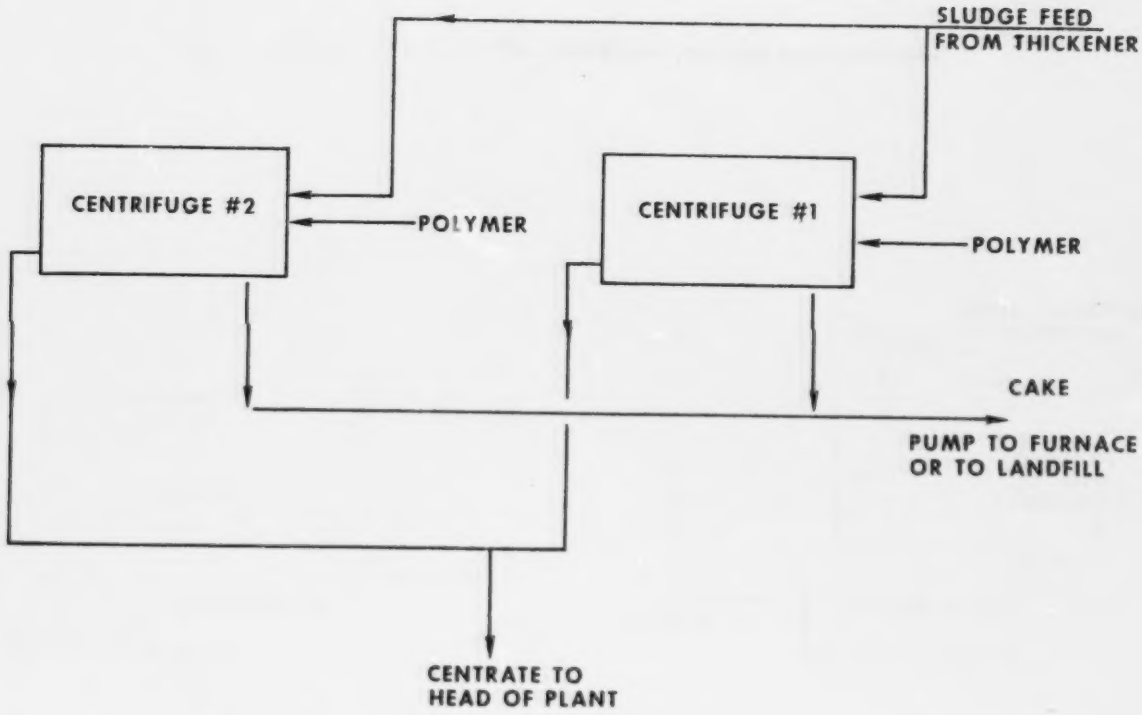

Figure 8. Centrifuge operation for total capture.

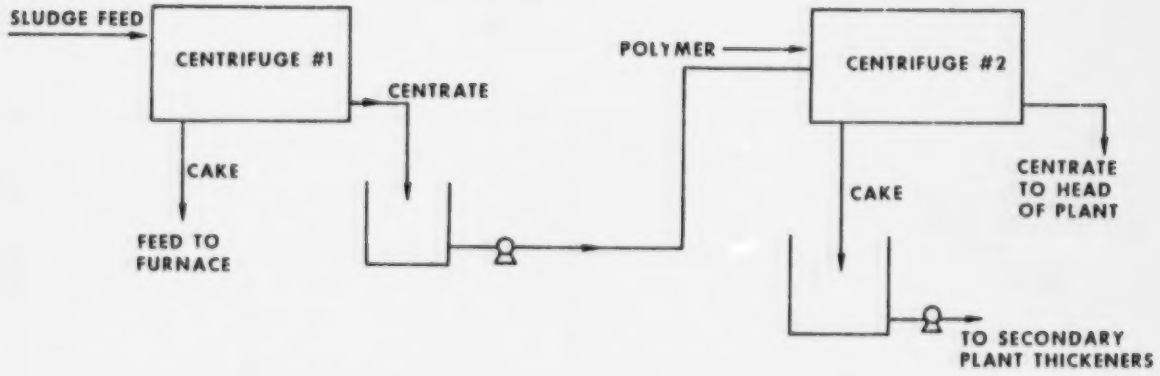

Figure 9. Centrifuge operation for wet classification.

During the classification operation, the cake from the second centrifuge, containing the inerts, was pumped to the secondary plant's thickeners. The cakes from the parallel centrifuges during total capture and from the first stage centrifuge during classification were carried by a conveyor belt to a progressive cavity pump. This wet cake was either pumped to the furnace for recalcination or drying, or pumped to a truck for land disposal.

An 18 ft diameter multiple hearth furnace was provided for recalcination of the chemical solids. The furnace is comprised of 6 circular hearths oriented horizontally and arranged for feeding solids at the top level. Hearths 1, 2 and 6 have four rabble arms each while hearths 3, 4 and 5 are provided with two arms per hearth. The rabble arms are equipped with teeth that are angled to rabble or rake the solids in a spiral motion across the hearths. As seen in Fig 10, the solids are directed on the first hearth from the outside of the furnace toward the center shaft where the solids drop onto the second hearth. The procedure is repeated through the six hearths with the solids reaching a maximum temperature of 1850°F.

The moisture is evaporated from the lime sludge by hot combustion gases which have been released from the auxiliary fuel burners mounted in hearths 2, 3, 4 and 5. The burners are supplied with #2 fuel oil that is mixed with air supplied by a 20 hp combustion blower. The blowers are angled to produce a circular motion of air and prevent the direct contact of the flame with the furnace components. The pilot lights for each burner are fired by natural gas. An automatic temperature control system is included.

The center shaft and rabble arms are cooled by a steady flow of air supplied by a 10 hp fan motor. The exhaust gases are drawn off by a 100 hp induced draft fan. The gases are first processed through a venturi scrubber then pass into an entrainment separator. A flow of water of 164 gpm was required by the venturi scrubber. A portion of the flue gases is compressed to 8 psig and piped into the liquid treatment system for pH control.

The recalcined solids discharge from the sixth hearth to the thermal disc product cooler at approximately 1600°F. The cooler includes three rows of hollow water-cooled metal discs that rotate through the lime as it passes through the cooler. The system is designed to cool the material from 1600°F at a maximum capacity of 1520 lb/hr. The cooling water is combined with water flowing to the scrubber and is included in the total 164 gpm water flow through the scrubber.

The cooled lime passes through the lump breaker and into an air-tight rotary lock feeder which transports the material to the recalcination storage bin by the introduction of pressurized air. The bin is equipped with an air-powered mechanical hammer which prevents bridging. A bag type dust collection system is also included. The recalcined lime is fed to the gravimetric feeder by a twin screw feeder.

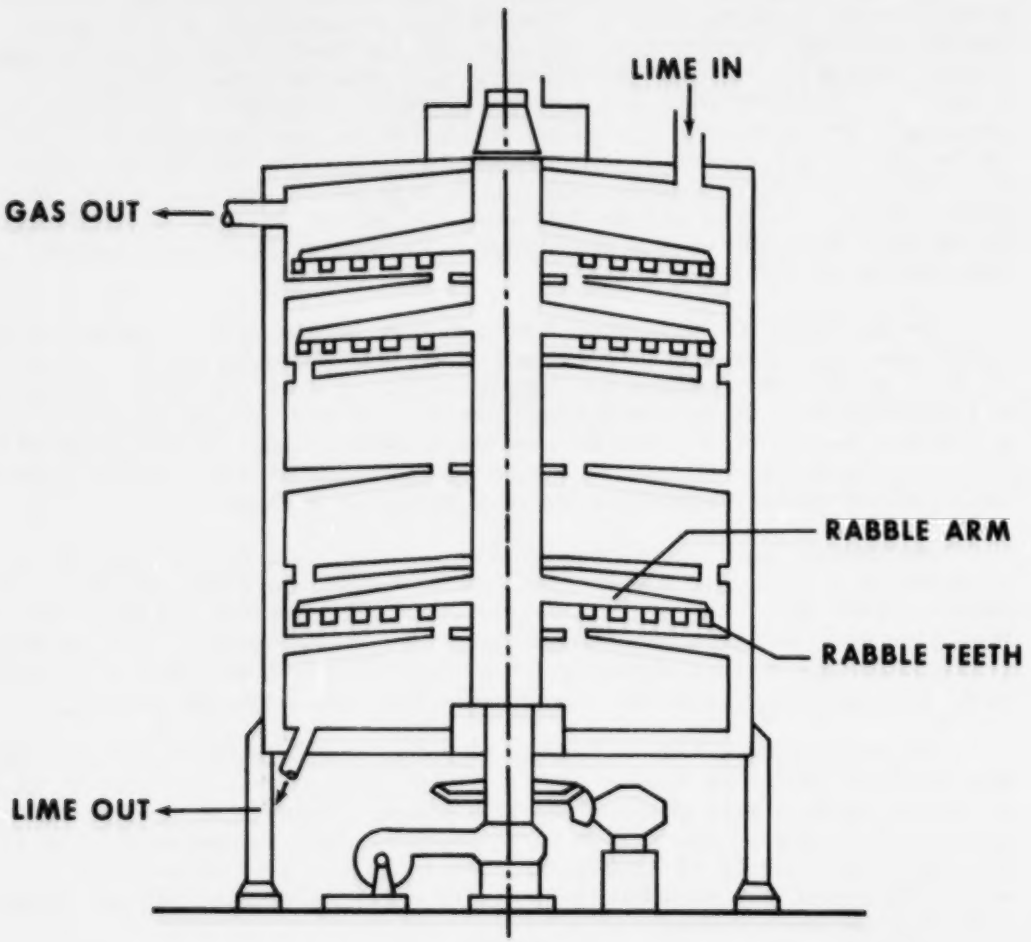

Figure 10. Cross section of multiple hearth furnace.

Filtration

The filter inlet well tank was located outside the operations building and collected flow from the second stage clarifier during the two-stage operation and from the recarbonation basin during the single stage operation. The purpose of the tank was to provide level control for the gavity filters. The inlet well was equipped with a 40 hp turbine mixer and a 6 in flue gas line for final pH adjustment prior to filtration. The feed system was made oversize as a safety factor to prevent calcium carbonate scaling of the dual media filters if the primary recarbonation system failed to produce the proper pH.

From the inlet well the flow passes into the operations building and is fed to five of the six dual media filters. Each filter has a surface area of 242 sq ft with a hydraulic loading of 3.1 gpm/sq ft at 1 mgd/filter. A side view of a dual media filter is shown in Fig. 11. Normally, the sixth filter is on standby and ready to be placed in service when required. The filter media consists of the following:

Material	Depth	Size
Anthracite Coal	18 inches	0.85-0.95 mm
Sand	6 inches	0.40-0.45 mm
Support Gravel	3 Inches	10 mesh - 3/16 in
	3 inches	3/16 in - 3/8 in
	3 inches	3/8 in - 5/8 in
	3 inches	5/8 in - 1 in

The flow through the filters is controlled by valves located at the discharge of each unit. Signals from the main control panel network control the five operating valves to maintain an even 5-way split of flow and a preset water level. Once a control valve reaches 100% open and thus can no longer control the flow, the filter is taken out of service and backwashed. However, for ease of scheduling, the units are backwashed every 24 hours at approximately 8 ft of headloss. A total of 12 ft of available head was included in the design of the units.

The dual media filters are cleaned in 10 minutes by the combination of surface wash and vertical backwash streams. Once the operator initiates the backwash cycle the procedure continues automatically. The sequence includes the following:

1. The units are drained to the drain troughs.
2. Backwash for 2 minutes at 10 gpm/sq ft.

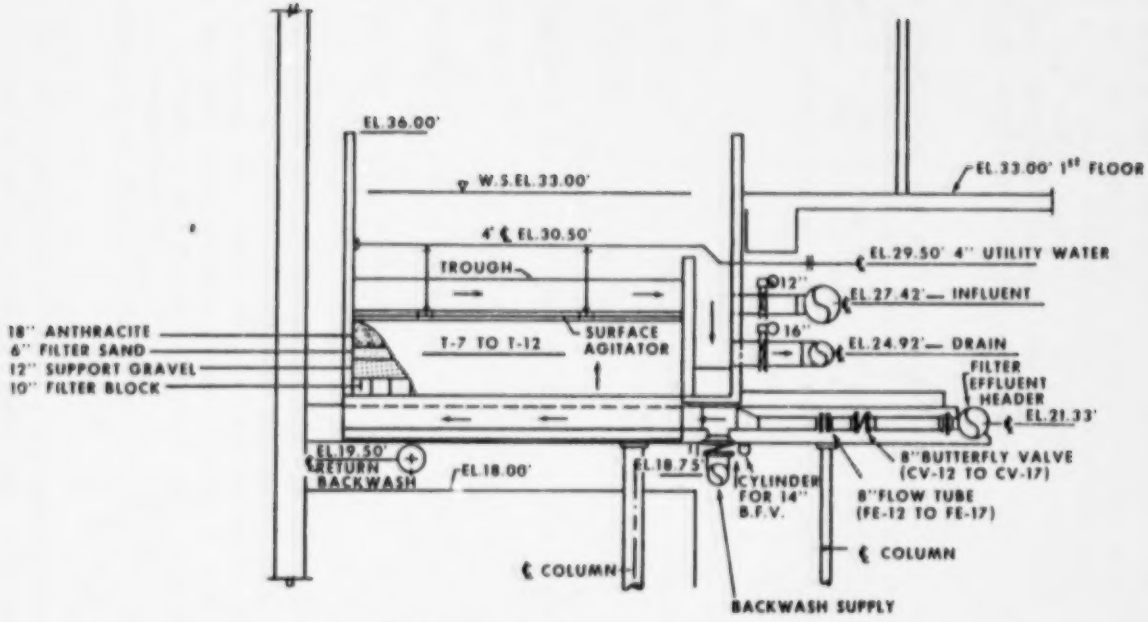

Figure 11. Cross section of dual media filter.

3. Flow is increased to 20 gpm/sq ft for 6 minutes for cleaning of the filter media.

4. The flow is decreased to 10 gpm/sq ft for 2 minutes to allow uniform settling of the filter media.

5. The hydraulically driven surface wash mechanisms are in service until the final 2 minutes.

The water supply for the backwash operation is filter effluent stored in a 59,000 gallon tank (equivalent to 1.8 backwash) located below the filter. Two backwash pumps are provided, each rated at 5000 gpm at 80 ft of total dynamic head. The backwash water from the filter is collected in another 59,000 gallon tank and returned to the head of the plant at the inlet structure at a rate of 250 gpm. The reduced flow is provided to prevent shock loads, both hydraulic and solids, to the first clarifier. The water for the surface wash mechanism is carbon adsorber effluent.

The effluents from the operating filters are collected in a common line and pass into the stabilization basin where flue gas is used to further adjust the pH, if necessary, prior to carbon adsorption. The 34 ft x 17 ft x 12 ft tank is covered and vented to prevent discharge of the toxic flue gas within the operations building. The unit also provides flow equalization ahead of the carbon adsorbers. Two 15 hp turbine mixers with two 4 in flue gas lines are installed. The effluent from the stabilization basin passes through over-under baffles and into the carbon adsorption wet well.

Carbon Adsorption

The carbon adsorption system consists of six downflow pressurized vessels arranged in sets of two to provide three parallel trains of two columns each. At a flow of 5 mgd, the loading to each column is 6.5 gpm/sq ft. A depth of 16 ft of activated carbon provides 18 minutes of Empty Bed Contact Time (EBCT) per column. The granular activated carbon was Filtrasorb 300* (8 x 30 mesh). The gravel support media for the activated carbon was similar to that provided in the dual media filters. Each of the three carbon trains is provided with a 50 hp, 2000 gpm centrifugal pump. The present flow control loop used to split and control the flow through the carbon adsorption system is similar to the system provided in the filters. The level in the carbon column inlet well was maintained by 3 control valves placed at the ends of the 3 parallel carbon trains. The control valves compensate for changing headlosses and maintain an even 3-way split of the flow and steady level in the inlet well by an analog logic network located in the main control panel.

The lead carbon column of each train was backwashed daily with the final column being backwashed every other day. The backwash cycle is initiated by the operator but then proceeds automatically. A series of on-off valves automatically isolate the columns to be backwashed, thus during

*A product of Calgon Corp., Pittsburgh, PA.

backwash, for approximately 10 minutes, one-third of the flow receives the treatment of only one carbon column. The time intervals of backwash include 2 minutes at 8.5 gpm/sq ft, 6 minutes at 17.5 gpm/sq ft and 2 minutes at 8.5 gpm/sq ft. A hydraulic surface wash mechanism is located 6 inches above the top of the carbon bed. The filters and carbon columns utilize the common 59,000 gallon backwash water supply tank. The backwash water flows through the column into the carbon adsorption sump tank from which it is returned to the head of the Model Plant at a rate of 250 gpm.

The carbon regeneration system is shown in Fig. 12. The exhausted carbon is removed from the activated carbon column through the four funnels located in the underdrain of each column (Fig. 13) and hydraulically carried to the exhausted carbon storage tank. The storage tank is equipped with a 1 in eductor to feed the carbon to the dewatering screw at a rate of 4750 lb/day. Initially, the transfer line was 1 in fiberglass with 90° elbows but was later converted to 1-1/4 in flexible tubing to prevent plugging. The concentration of carbon in the slurry is approximately 1 lb/gallon. The carbon slurry is dewatered in an inclined screw conveyor.

The dewatered carbon is fed to the second hearth of the 54-in diameter multiple hearth furnace. The carbon furnace, like the lime recalcination furnace, is a multiple hearth. The unit has five hearths with rabble arms on four levels to move the carbon downward through the hearths. The top hearth is used as an afterburner for the ignition of the off gases from the carbon regeneration process. The exhaust gases pass through a venturi scrubber and entrainment separator. The wash water from the separator is returned to the model plant's sump.

The carbon passes through the four lower hearths and discharges to the quench tank where the carbon is rapidly cooled in water. Two of the four hearths which receive carbon are equipped with burners which are fired by natural gas. A blower supplies a steady flow of air to cool the center shaft and rabble arms. From the quench tank, the activated carbon is hydraulically transported to the regenerated carbon storage tank where it is held until the regeneration of another column is required.

The effluents from the three carbon systems are collected in a common line and flow to the polishing ponds. The effluent from the ponds receiving carbon effluents is combined with the effluents from the ponds containing Piscataway secondary effluent, chlorinated and discharged to Piscataway Bay.

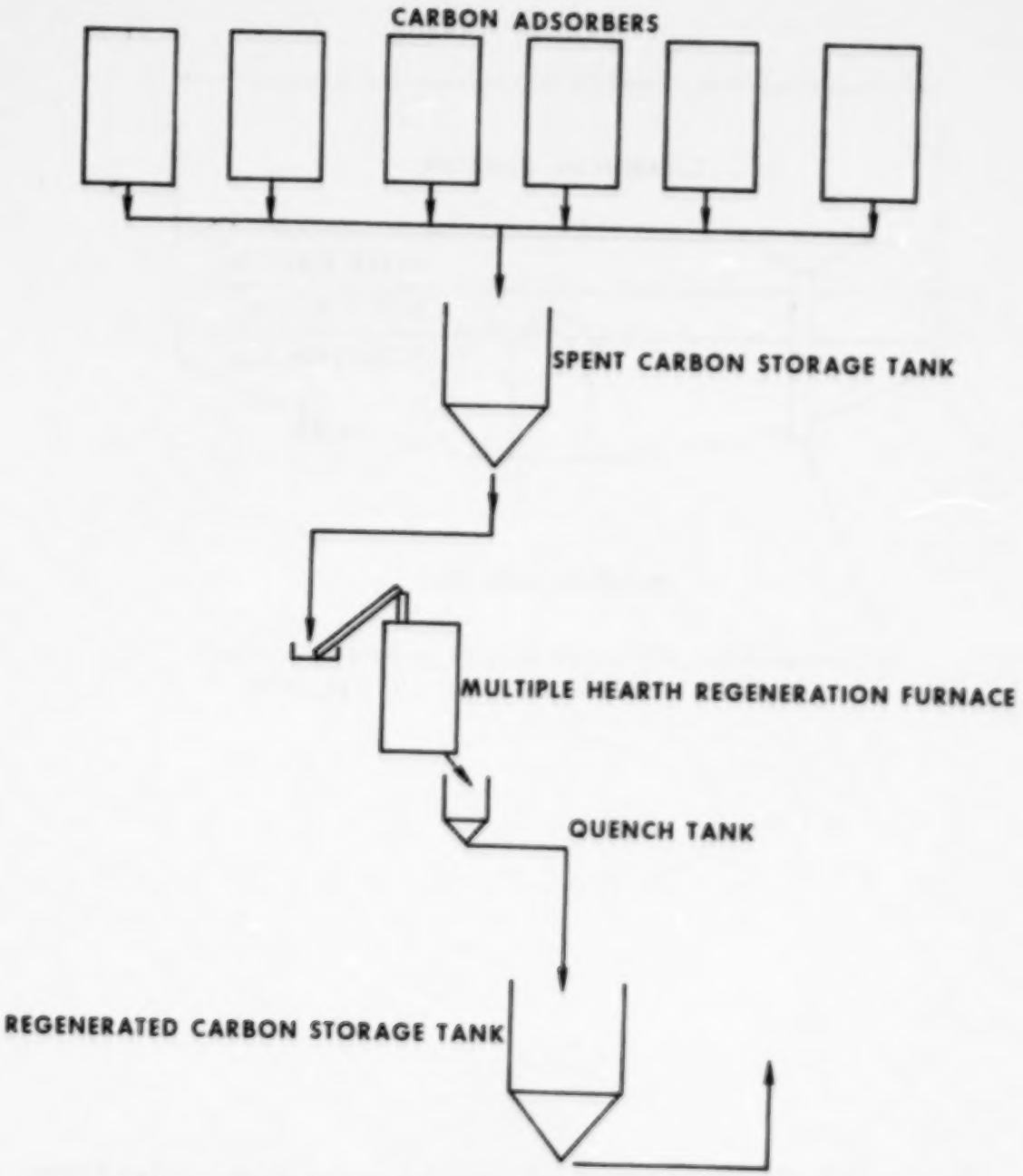

Figure 12. Flow schematic for carbon regeneration.

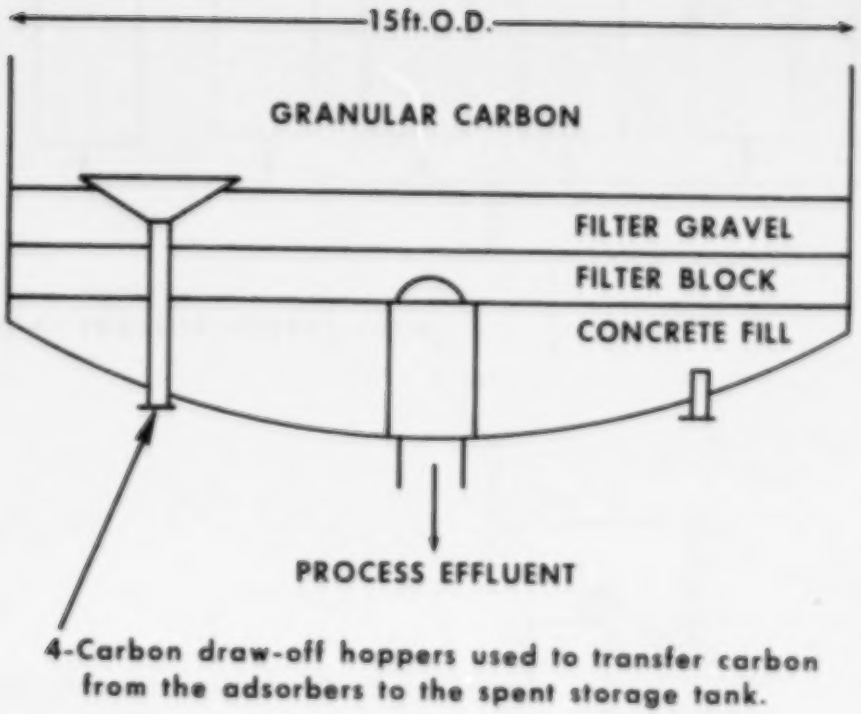

Figure 13. Cross section of carbon adsorber underdrain.

VI. RESULTS OF TWO-STAGE HIGH LIME EVALUATION

The operation of the two-stage high lime system was sustained at the design flow rate for 36 days in October and November, 1973, a period between two failures of the reactor clarifiers.

Secondary Operation

The operation of the step aeration activated sludge system described in Table 4 corresponds to the 36 days of operation of the tertiary facility and was typical of that achieved during the two years of the grant period. The mixed liquor suspended solids concentration of 2133 mg/l is an average of samples taken at the quarter points along the reactor. As seen by the average decrease in alkalinity from 135 to 97 mg/l as $CaCO_3$, the activated sludge system was nitrifying. The reduction in alkalinity is a result of the production of nitric acid during nitrification. The variations in alkalinity and TKN in the secondary effluent are shown in Fig. 14. The degree of nitrification was inconsistent, most probably because of the irregular wasting schedule from the secondary settlers. The SRT of 6.6 days is based on the average wasting rate for the 36 day operating period. The average concentrations of BOD and suspended solids for the raw wastewater, primary effluent and secondary effluent are presented in Tables 5 and 6. The effect of the recycle of solids to the grit chamber is reflected in the high values for these parameters in the primary effluent.

The phosphorus concentrations are given in Table 7. During the evaluation period for the tertiary lime treatment systems, 80 mg/l alum was being added to the aeration tanks of the 25 mgd activated sludge system to improve removals of phosphorus and suspended solids. Return streams to the head of the plant containing high concentrations of alum sludge and precipitated phosphate entered the 5 mgd secondary system and produced high phosphorus concentrations in the primary effluent. The overall effect of the return stream was an increase in phosphorus removal through the old 5 mgd secondary plant. During the 36 days of operation, the secondary plant removed 55.7% of the incoming phosphorus. The average nitrogen concentrations are given in Table 8, but do not reflect the variability as seen in Fig. 14.

Tertiary Treatment

During the 36 days of continuous operation the influent flow to the Model Tertiary Plant averaged 4.586 mgd. The hydraulic loadings to the unit processes based on influent flow plus recycle are presented in Table 9. The recycled plant water was 11.3% of the total flow. A summary of the plant recycle flows is presented in Table 10.

TABLE 4

OPERATING CONDITIONS OF THE PISCATAWAY SECONDARY PLANT
DURING THE HIGH LIME PROCESS EVALUATION

DAILY FLOW, mgd	5.675
DETENTION TIME, hr	4.4
MLSS, mg/l	2133
RECYCLE RATE, %	37
SOLIDS IN RECYCLE, mg/l	7970
WASTE RATE, 1000 gal/day	38.1
SVI, ml/gm	96
SRT, days	6.6
F/M, lb BOD_5/lb MLVSS	0.41
RAW WASTEWATER pH	7.2
SECONDARY EFFLUENT pH	7.4
RAW WASTEWATER TEMPERATURE, °F	65
RAW WASTEWATER ALKALINITY, mg/l $CaCO_3$	135
SECONDARY EFFLUENT ALKALINITY, mg/l $CaCO_3$	97

TABLE 5

REMOVAL OF BIOCHEMICAL OXYGEN DEMAND (BOD 5 DAY)
DURING THE HIGH LIME PROCESS EVALUATION

	mg/l	% Removal
Raw	141.0	---
Primary	145.0	---
Secondary	16.5	88.3
Lime clarified	5.9	95.8
Filtered	5.7	96.0
Carbon Adsorption	4.0	97.2

Note: Secondary Plant recycle enters between the raw sample point and the primary clarifier.

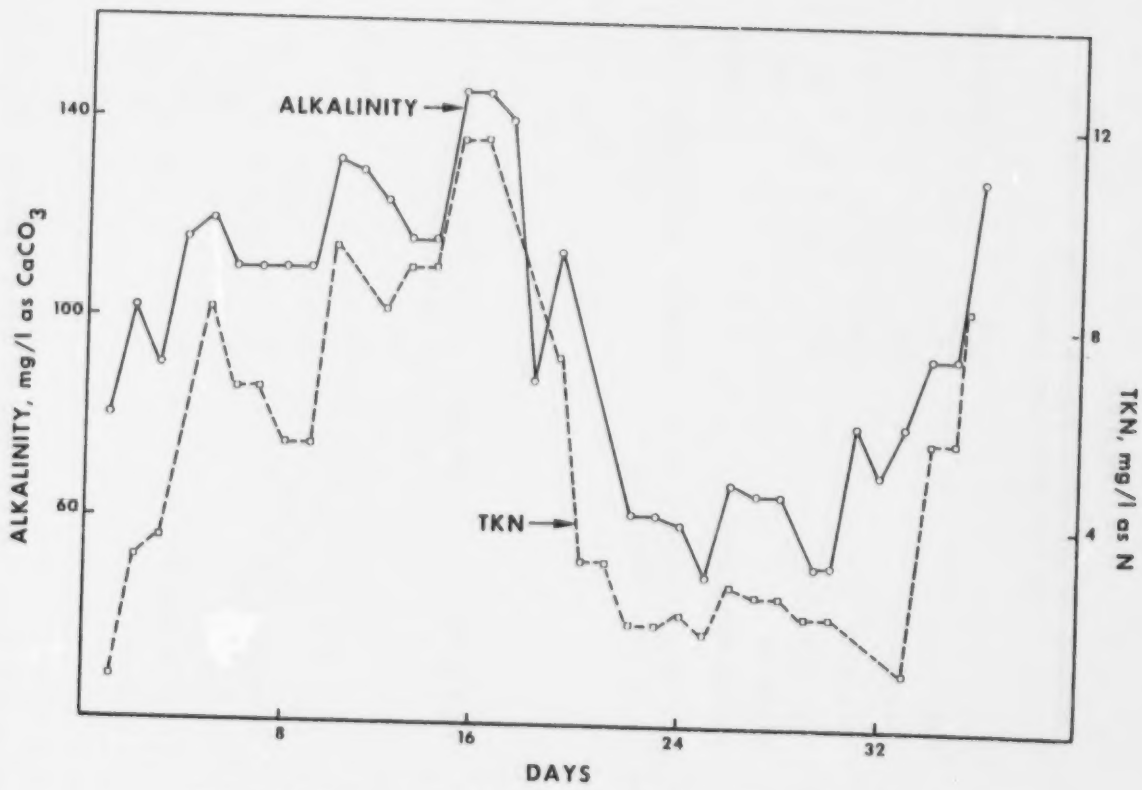

Figure 14. Comparison of alkalinity and TKN of secondary effluent.

TABLE 6

REMOVAL OF SUSPENDED SOLIDS DURING EVALUATION
OF THE HIGH LIME PROCESS

	mg/l	% Removal
Raw	121	---
Primary	183	---
Secondary	27.5	77.3
Lime Clarified	21	82.6
Filtered	6	95.0
Carbon Adsorption	2.5	97.9

Note: Secondary Plant recycle enters between the raw sample point and the primary clarifier.

TABLE 7

REMOVAL OF TOTAL PHOSPHORUS (AS P) DURING EVALUATION
OF THE HIGH LIME PROCESS

	mg/l	% Removal
Raw	7.90	---
Primary	9.60	---
Secondary	3.50	55.7
Lime Clarified	0.26	96.7
Filtered	0.20	97.5
Carbon Adsorption	0.10	98.7

Note:
Secondary Plant recycle enters between the raw sample point and the primary clarifier

TABLE 8

REMOVAL OF NITROGEN* COMPOUNDS DURING EVALUATION OF THE HIGH LIME PROCESS

	NH_3	TKN	NO_3	NO_2	TOTAL N
Raw	13.1	16.2	1.1	0.1	17.4
Primary	---	---	---	---	---
Secondary	4.5	5.2	6.7	.4	12.3
Lime Clarified	5.4	6.0	6.4	.3	12.7
Filtered	4.1	---	5.9	---	---
Carbon Adsorption	2.9	3.3	8.1	.1	11.5

*All values mg/l as N

TABLE 9

LOADING RATES DURING EVALUATION OF THE HIGH LIME PROCESS

Secondary Plant

Flow 5.7 mgd

Primary Clarifiers	2-60 ft dia.	1008 gpd/sq ft
Secondary Clarifiers	2-70 ft dia.	741 gpd/sq ft

Model AWT Plant

External Flow and Recycle = Total Flow
4.586 mgd + 0.588 mgd = 5.174 mgd

First Stage Reactor Clarifier 1-80 ft dia.	1029 gpd/sq ft
Second Stage Reactor Clarifier 1-70 ft dia.	1344 gpd/sq ft
Dual Media Filtration (5 units) 242 sq ft/unit	4276 gpd/sq ft or 2.97 gpm/sq ft

Carbon Adsorption

Column Set #1 Avg. Flow 2.063 mgd	8.1 gpm/sq ft
Column Set #2 Avg. Flow 1.420 mgd	5.6 gpm/sq ft
Column Set #3 Avg. Flow 1.309 mgd	5.1 gpm/sq ft

All loadings are based on an average flow of 5.174 mgd.

TABLE 10

PLANT RECYCLE FLOWS DURING HIGH LIME EVALUATION

	SOURCES	TOTAL GALLONS/DAY	% OF FLOW
1	FILTER BACKWASH		
	38,720 gal x 6 filters/day	232,320	4.49
2	CARBON COLUMN BACKWASH		
	24,017 gal x 4 columns/day	96,068	1.86
3	RECALCINATION FURNACE		
	164 gal/min x 1440 min/day	236,160	4.56
4	Misc. - (centrate, pump sealing		
	Water, flushing & wash water)	23,452	0.45
	TOTAL	588,000	11.36

All percentages are based on an average flow of 5.174 mgd.

The first reactor clarifier was maintained at pH 11.45 by an average lime dose of 257 mg/l as CaO. The lime dose calculation is based on total pounds of recalcined and virgin pebble limes added to the system with average Available Lime Indexes (ALI) of 60 and 87%, respectively. The quantities of chemicals added to the system during the 36 days of operation are presented in Table 11. During the operation of the two stage system, the solids handling system was operated for total capture with wasting of the recalcined lime to prevent build-up of inerts in the system. As a result, the low ALI of 60% was produced. The recalcined lime accounted for 75% of the total material added to the system, or 68% of the required calcium oxide.

The effluent from the first clarifier was reduced to an average pH 10 in the recarbonation basin. During periods of zero or low sludge feed to the furnace, the percent of carbon dioxide in the flue gas decreased to approximately 5% which was insufficient to maintain the proper pH. With adequate sludge feed to the furnace the carbon dioxide concentration was approximately 12% which was sufficient for pH control. Additional carbon dioxide feed from the liquid storage tank was required on 14 of the 36 operating days. Because of the difficulty in maintaining proper control of the 98% and 5% carbon dioxide feed systems, an accurate measurement of the amounts of carbon dioxide added to the system, was not possible. Sludge from the second clarifier was recycled to the recarbonation basin at a rate of 5% of the average daily flow.

Following recarbonation, the effluent was fed to the second stage clarifier for settling of the precipitated calcium carbonate. A ferric chloride solution at an average concentration of 17.8 mg/l as $FeCl_3$ was added to the draft tube of the second unit to improve clarification. The waste sludge from the second clarifier was pumped to the first clarifier's draft tube at a rate of 0.66% of average daily flow. Total solids wasting from the system was accomplished by pumping the sludge from the first clarifier to the gravity thickener at an average pumping rate of 0.7% of average flow.

The results of the operation of the chemical clarification system are presented in Tables 5 through 8 and in Table 12. As expected, the lime clarification system reduced the phosphorus concentration to low levels, 0.26 mg/l as P. With efficient insolubilization of the BOD in the secondary treatment system and capture of the secondary effluent suspended solids in the chemical clarification system, an average BOD concentration of 5.9 mg/l was produced from the lime treatment system.

The second stage effluent was reduced to pH 8 by the addition of flue gas and supplemental carbon dioxide in the filter inlet well prior to dual media filtration. As with the recarbonation basin, an accurate measurement of the amount of carbon dioxide could not be determined.

Five dual media filters were used continuously with another filter either in backwash or standby. The flow rate of the filters averaged 2.97 gpm/sq ft. The filters were backwashed on a 24-hour cycle for ease of operation at an average head loss of 8 ft.

TABLE 11

CHEMICAL USAGE IN THE HIGH LIME PROCESS

Lime	
Total Virgin Pounds	137,079
Average Available Lime Index (ALI), %	87
Available CaO, lb	119,259
Average Daily Usage, lb	3,138
Average Daily Dose, mg/l	87
Total Recalcined Pounds	426,618
Average Available Lime Index (ALI),%	60
Available CaO, lb	255,971
Average Daily Usage, lb	6,731
Average Daily Dose, mg/l	170
Ferric Chloride ($FeCl_3$)	
Total Pounds added to clarifier,lb	23,496
Average Daily Usage,lb	618
Average Daily Dose, mg/l	17.8
Polymer Usage (Centrifuge Only)	
Total Pounds used	568.3
Average Daily Usage, lb	15.0
Pounds Polymer/Ton of Dry Sludge	0.25

TABLE 12

REMOVALS OF CHEMICAL OXYGEN DEMAND (COD) AND TOTAL ORGANIC CARBON (TOC) DURING EVALUATION OF THE HIGH LIME PROCESS

	COD mg/l	TOC mg/l as C
Raw	---	---
Primary	---	---
Secondary	34.8	12.3
Clarified	---	13.4
Filtered	24.7	7.5
Carbon Adsorption	13.4	1.8

The dual media filters were not effective in removing additional materials with the exception of suspended solids. The calculated removal of suspended solids in the filters was 15 mg/l. However, since the suspended solids in the influent to the filter system consists mainly of precipitated calcium carbonate, a portion of which is solubilized in the filter inlet well, an accurate efficiency of the filters alone cannot be determined. However, if one assumes that 15 mg/l of suspended solids is captured in one 24-hour run at 1 mgd, an estimate of the efficiency of the filters can be determined. A total of 125 lb of solids would be captured in each filter for a loading of 0.52 lb/sq ft/cycle. By using a termination headloss of 8 ft, then the loading can be expressed as 0.065 lb/sq ft/ft of headloss. These loadings are quite reasonable considering that the units contain 24 inches of filter media compared to 36 inches of media in most designs.

The filter effluent was reduced to pH 7.2 by the addition of flue gas in the stabilization basin and pumped to the three parallel activated carbon adsorption systems. As seen in Table 9, the hydraulic loadings to the three systems averaged 8.1, 5.6 and 5.1 gpm/sq ft. The inconsistency in the loadings was due to the inability of the control system to maintain an even split in flow. The lead column of each system was backwashed every 24 hours with the final column being backwashed every 48 hours.

Approximately 24,000 gallons of water was required to backwash a carbon column as compared to 39,000 gallons to backwash a dual media filter. During the backwash of a column, treatment by only one column in that train was provided.

The performance of the activated carbon system based on composite samples from the three trains is presented in Tables 5 through 8. The results are the averages for the 36 days of operation. As seen in Table 12, the units removed approximately 75% of the inlet TOC and 50% of the COD which is typical of activated carbon operation at other locations. The BOD analysis was uninhibited and included the effects of sample nitrification in addition to oxidation of organics. The degree of nitrification in the activated carbon columns, from 5.9 to 8.1 mg/l of NO_3-N as shown in Table 8, was typical of the operation of the units throughout the study.

The performances of the individual columns during the 36-day high lime evaluation are presented in Tables 13, 14 and 15. The cumulative loadings to the carbon columns, based on the cumulative flow through each column from March 1973 through November 1973 are presented in Table 16. Column T-16 was shut down for nearly two months prior to October 13, 1973, because of a broken seat in the discharge valve during which period only T-17 was in operation. At reduced flows, only columns T-14 and 15 were in operation which accounts for their higher loadings and cumulative flows. BOD loadings are not presented because of the effect of nitrification on the BOD analysis.

With the failure of the chemical clarifiers on November 18, 1973, carbon column T-14 was prepared for regeneration. As seen in Tables 13 and 16, the carbon was not exhausted but was still efficient in removing organics. It was obvious, however, that the plant was to be down for an extended period and the information and experience obtained from a regeneration cycle within the grant period was considered to be important. In addition, with the remainder of the plant shut down, full operator attention could be focused on the carbon regeneration system. The results of the carbon regeneration operation are presented in detail in Section VII.

The waste solids from the first clarifier, which included the solids wasted from the second clarifier, were pumped to the gravity thickener for solids processing. The solids concentration in the underflow from the first clarifier generally ranged from 7 to 10% and varied as a function of the pumping rate, which during the 36 days of operation averaged 0.70% of the influent flow or 32,000 gallons/day.

Three methods were used to determine the material balances around the clarification and solids handling system using the following data:

1. Chemical analyses of the liquid streams and measured daily influent flow rates.

2. Chemical analyses of the sludge streams and measured sludge flows.

3. Total solids concentrations and measured sludge flows.

TABLE 13

PERFORMANCE OF CARBON ADSORBER TRAIN #1 DURING
EVALUATION OF THE HIGH LIME PROCESS

Columns T-14 and T-15

Average Flow 2.063 mgd

Cumulative flow from March 1973 - November 1973 199.907 mil gal

Total Organic Carbon (TOC)	mg/l as C	% Removed
Influent	7.5	---
Intermediate	3.2	57.3
Effluent	2.3	69.3
Biochemical Oxygen Demand (BOD)	**mg/l**	
Influent	6.2	---
Intermediate	5.8	6.4
Effluent	4.0	35.5
Chemical Oxygen Demand (COD)	**mg/l**	
Influent	24.9	---
Intermediate	17.4	30.1
Effluent	15.1	39.4

Organic concentrations based on data from October 14 to November 18, 1973.

TABLE 14

PERFORMANCE OF CARBON ADSORBER TRAIN #2 DURING
EVALUATION OF THE HIGH LIME PROCESS

Columns T-16 and T-17

Average Flow 1.420 mgd

Cumulative flow from March 1973 - November 1973 173.151 mil gal

Total Organic Carbon (TOC)	mg/l as C	% Removed
Influent	6.5	---
Intermediate	---	---
Effluent	2.4	63.1
Biochemical Oxygen Demand (BOD)	**mg/l**	
Influent	6.0	---
Intermediate	---	---
Effluent	4.6	23.3
Chemical Oxygen Demand (COD)	**mg/l**	
Influent	24.2	---
Intermediate	---	---
Effluent	13.7	43.4

Organic concentrations based on data from October 14 to November 18, 1973.

TABLE 15

PERFORMANCE OF CARBON ADSORBER TRAIN #3 DURING
EVALUATION OF THE HIGH LIME PROCESS

Columns T-18 and T-19

Average Flow 1.309 mgd

Cumulative flow from March 1973 - November 1973 126.396 mil gal

Total Organic Carbon (TOC)	mg/l as C	% Removed
Influent	7.6	---
Intermediate	2.6	65.8
Effluent	1.5	80.3
Biochemical Oxygen Demand (BOD)	**mg/l**	
Influent	6.3	---
Intermediate	5.6	11.1
Effluent	4.5	28.6
Chemical Oxygen Demand (COD)	**mg/l**	
Influent	25.4	---
Intermediate	15.4	39.4
Effluent	12.4	51.2

Organic concentration based on data from October 14 to November 18, 1973.

TABLE 16

CUMULATIVE COD AND TOC LOADINGS ON CARBON AT THE END OF THE 36-DAY HIGH LIME EVALUATION

		lb COD/lb Carbon	Cumulative Flow, mil gal	lb TOC/lb Carbon
Set 1				
	T-14	0.21155	199.907	0.09613
	T-15	0.06893	199.907	0.01812
Set 2				
	T-16	0.09855	132.801	0.05230
	T-17	0.09553	173.151	0.03419
Set 3				
	T-18	0.12065	126.396	0.06207
	T-19	0.04086	126.396	0.01222

TABLE 17

SOLIDS MATERIAL BALANCES FOR THE HIGH LIME EVALUATION

FIRST CLARIFIER	
Solids Captured	17,400 lb/day
Solids Wasted to Thickener	22,400 lb/day
SECOND CLARIFIER	
Solids Captured	5,100 lb/day
Solids Wasted to First Clarifier	11,800 lb/day
TOTAL SOLIDS REPORTING TO FIRST CLARIFIER	22,500 lb/day
SOLIDS REPORTING TO CENTRIFUGE	24,700 lb/day
*SOLIDS REPORTING TO FURNACE	20,200 lb/day

*Measured by gravimetric feeder following recalcination furnace assuming a 75% loss through furnace.

The accuracy of method two is limited because of the analytical and sampling errors associated with the highly concentrated sludge streams. Method three was based on average solids concentrations which in practice were quite variable and therefore difficult to sample. The data for the liquid streams were considered to be the most reliable because of the accuracy of the chemical analyses in combination with good flow measurements.

Accepting these limitations, material balances around the clarifiers were calculated and are presented in Table 17. The average solids captured in the first and second stage clarifiers of 17,400 and 5,100 lb/day, respectively, are based on chemical analyses and flow measurements of the liquid streams. The inert solids in the recalcined lime and return flow (filter and carbon column backwash, centrate and scrubber) water are included in the solids captured in the first clarifier. The solids wasted to the first clarifier and thickener are based on average daily waste rates and total solids concentrations. The amount which appears to be in gross error is the 11,800 lb/day that reported from the second clarifier to the first clarifier. The concentration of total solids in the waste sludge was variable and no doubt resulted in an erroneous average number. By using the liquid stream approach then, as seen in Table 17, the amount of solids (22,500) that reported to the clarifier is close to the amount of solids that was pumped from the first clarifier to the thickener (22,400). The amount that reported to the thickener was based on measured total solids and waste flow rates. The data show that the combination of clarifier-thickener was in solids balance.

During the operation of the recalcination furnace, the underflow from the gravity thickener, containing approximately 20% total solids, was pumped to one of the two solid bowl centrifuges for dewatering. A moderately anionic polymer was added to the centrifuge at a rate of 0.61 lb polymer/ton dry solids for total capture. Sludge cake with a suspended solids concentration of 38.5% was produced and pumped to the top of the recalcination furnace.

The combination of centrifugation and recalcination was not operated continuously during the 36 days of high lime evaluation. The total solids production of 22,500 lb/day in the clarifiers was less than the capacity of the furnace. Since the solids handling system required high manpower, it was operated only to produce sufficient recalcined lime and/or maintain the sludge blanket level in the gravity thickener.

A total of 545,818 lb of recalcined lime was produced at an average available lime index of 60%. Since the centrifuge was operated for total capture of the solids, wasting of the recalcined lime to a sanitary landfill was required to prevent the buildup of inerts in the lime. Based on preliminary calculations, an estimated waste rate of 25% was established. A total of 144,200 lb of recalcined lime, or 25% of the total production, was wasted. The recalcined and pebble limes were mixed at a 75/25% ratio by two gravimetric feeders prior to slaking. With the exception of increased grit production, the paste slaker operation was reasonably successful.

A heat balance of the recalcination furnace was calculated based on average concentrations of materials in the centrifuge cake and average operating temperatures. The results of the heat balance are presented in the Appendix. The feed to the furnace averaged 3315 lb of wet sludge/hr or 1276 dry lb/hr.

In order to determine if the lime recovery system could supply on a daily basis, the recalcined lime actually added to the system, the furnace was assumed to be operated at 1276 dry lb/hr until the daily sludge production was recalcined. The length of time for furnace operations, based on a daily sludge production of 22,500 dry lb/day, was determined to be 17.6 hr. The total amount of CaO produced in the furnace, based on the data in the Appendix, would equal 9046 lb/day (514.6 lb/hr x 17.6 hr). Assuming a waste rate of 26% or 2351 lb/day, the total calcium oxide available for reuse would equal 6695 lb/day. As seen in Table 11, the average daily usage of CaO was actually 6731 lb/day. Based on this information it can be assumed that the clarification and recalcination systems were in reasonable balance.

The heat balance around the furnace was good with 94% of the fuel used accounted for in the calculation. The importance of reducing the moisture concentration of the sludge feed should be noted. The BTU requirement for the evaporation of the moisture accounted for 58% of the total BTU input.

VII. RESULTS OF SINGLE-STAGE LOW LIME EVALUATION

The second major operational system to be tested was single-stage low lime. This system was started on April 10, 1974, following replacement of main gears in both clarifiers, and operated for 89 days until July 8, 1974, when the induced draft fan on the recalcination furnace malfunctioned.

Secondary Operation

The operation of the secondary facility was generally typical of that described for the two-stage high lime mode. Several factors were different, however, as shown in Table 18 which represents the operational results for the 89-day operating period. The mixed liquor suspended solids concentration of 2600 mg/l is an average of four samples taken at the quarter points along the reactor. The major differences in the secondary operations for the high lime and low lime operations were in the sludge retention times and the recycle rates. As seen by the decrease in alkalinity from 145 to 119 mg/l, the activated sludge system was nitrifying. The waste rate of 50,280 gal/day was the average rate for the total period. Because of the configurations of the secondary facility and the two separate systems used, wasting from the aeration basins was conducted as a batch operation.

The average concentrations of BOD and suspended solids for the raw wastewater, primary effluent and secondary effluent are presented in Tables 19 and 20. The effects of the recycle of solids to the grit chamber are reflected in the high values for the primary effluent. The recycled solids to the grit chamber are from the overloaded thickeners and solids handling system. The phosphorus concentrations, as given in Table 21, again show the results of recycling the aluminum phosphate enriched flow to the grit chambers. A secondary plant removal of 58.5% of the raw sewage phosphorus was obtained. The nitrogen concentrations are given in Table 22.

Tertiary Treatment

The single stage low lime tertiary system as shown in Fig. 4, operated at an average daily flow of 4.092 mgd. The hydraulic loadings to the unit processes based on influent flows plus recycle are presented in Table 23. As shown in Table 24, the recycle of the plant water accounted for 12% of the total flow in the tertiary system.

The reactor clarifier was maintained at pH 10.4 by an average lime dose of 113.4 mg/l and a ferric chloride dose of 25.2 mg/l as $FeCl_3$ as shown in Table 25. All solids formed in the lime clarification system were precipitated in the reactor-clarifier and were pumped from the center collection hopper to the gravity thickener. As seen in Table 25, most of the lime used

TABLE 18

OPERATING CONDITIONS OF THE PISCATAWAY SECONDARY OPERATION DURING THE LOW LIME PROCESS EVALUATION

Daily Flow, mgd	5.276
Aeration Tank Detention Time, hr	3.0
MLSS (% Volatile), mg/l	2600 (64)
Recycle Rate, %	48
Solids in Recycle, mg/l	6922
Waste Rate, 1000 gal/day	50.28
SVI, ml/gm	104
SRT, days	2.67
F/M, lb BOD_5/lb MLVSS	0.47
Raw Wastewater, pH	7.0
Secondary Effluent, pH	7.3
Raw Wastewater Temperature, °F	62
Raw Wastewater Alkalinity, mg/l $CaCO_3$	145
Secondary Effluent Alkalinity, mg/l $CaCO_3$	119

TABLE 19

REMOVAL OF BIOCHEMICAL OXYGEN DEMAND (BOD 5 DAY) DURING EVALUATION OF THE LOW LIME PROCESS

	mg/l	% Removal
Raw	123.9	---
Primary	200.6	---
Secondary	16.0	87.1
Clarified	8.2	93.4
Filtered	6.6	94.7
Carbon Adsorption	2.3	98.2

Note:
Secondary Plant recycle enters between the raw sample point and the primary clarifier.

TABLE 20

REMOVAL OF SUSPENDED SOLIDS DURING EVALUATION OF THE LOW LIME PROCESS

	mg/l	% Removal
Raw	130.2	---
Primary	376.8	---
Secondary	18.5	85.8
Lime Clarified	15.8	87.9
Filtered	10.2	92.2
Carbon Adsorption	3.5	97.3

Note:

Secondary Plant recycle enters between the raw sample point and the primary clarifier.

TABLE 21

REMOVAL OF TOTAL PHOSPHORUS (as P) DURING EVALUATION OF THE LOW LIME PROCESS

	mg/l	% Removal
Raw	6.63	---
Primary	19.31	---
Secondary	2.75	58.5
Lime Clarified	.56	91.6
Filtered	.29	95.6
Carbon Adsorption	.16	97.6

Note:

Secondary Plant recycle enters between the raw sample point and the primary clarifier.

TABLE 22

REMOVAL OF NITROGEN COMPOUNDS* DURING
EVALUATION OF THE LOW LIME PROCESS

	NH_3	TKN	NO_3	NO_2	TOTAL N
Raw	---	---	---	---	---
Primary	---	---	---	---	---
Secondary	7.92	8.87	5.02	.69	14.59
Lime Clarified	9.64	10.37	3.67	.64	14.68
Filtered	9.11	9.92	4.38	.58	14.88
Carbon Adsorption	7.60	8.34	5.16	.48	13.98

* All values mg/l as N

TABLE 23

LOADING RATES DURING EVALUATION OF THE LOW LIME PROCESS

Secondary Plant		
Flow 4.858 mgd		
Primary Clarifiers	2-60 ft dia.	859 gpd/sq ft
Secondary Clarifiers	2-70 ft dia.	631 gpd/sq ft
Model AWT Plant*		
External Flow and Recycle = Total Flow		
4.092 mgd + 0.588 mgd = 4.680 mgd		
First Stage Reactor Clarifier	1-80 ft dia.	931 gpd/sq ft
Second Stage Reactor Clarifier	1-70 ft dia.	1216 gpd/sq ft
Dual Media Filtration (5 units) 242 sq ft/unit		3867 gpd/sq ft or 2.69 gpm/sq ft
Carbon Adsorption		
Column Set #1 Avg. Flow 1.285 mgd		5.05 gpm/sq ft
Column Set #2 Avg. Flow 1.117 mgd		4.39 gpm/sq ft
Column Set #3 Avg. Flow 1.095 mgd		4.31 gpm/sq ft

* Based on flow plus internal recycle streams.

TABLE 24

PLANT RECYCLE FLOWS DURING LOW LIME EVALUATION

Sources	Total gal/day	% of Flow
1. Filter Backwash 38,720 gal x 6 filters/day	232,320	4.96
2. Carbon Column Backwash 24,017 gal x 4 columns/day	96,068	2.05
3. Recalcination Furnace 164 gal/min x 1440 min/day	236,160	5.05
4. Misc. (centrate, pump sealing water, flushing and wash water)	23,452	0.50
Total	588,000	12.56

All percentages are based on an average flow of 4.680 mgd.

TABLE 25

CHEMICAL USAGE IN THE LOW LIME PROCESS

Lime	
Total Virgin Pounds	390,829
Average Available Lime Index (ALI)	87%
Available CaO, lb	340,018
Average Daily Usage, lb	4,097
Average dose, mg/l	105.3
Total Recalcined Pounds	37,088
Average Available Lime Index (ALI)	50.8%
Available CaO,lb	18,841
Average Daily Usage, lb	224
Average Dose, mg/l	8.1
Ferric Chloride ($FeCl_3$)	
Total Pounds added to clarifier	68,646
Average Daily Usage, lb	817
Average Daily Dose, mg/l	25.2
Polymer Usage (Centrifuge Only)	
Total pounds	1,828
Average Daily Usage, lb	26.1
Pounds Polymer/Ton of Dry Sludge	6.8

during the test period was virgin pebble lime with an available lime index of 87%. The precipitated lime was recalcined and recycled back into the system for only a short time. The recalcined lime had an average available lime index of only 50.8%. In the single stage low lime system, carbon dioxide is used to reduce the pH below the solubility point of calcium carbonate, thus preventing a lime scale problem in the filters. The pH reduction was accomplished in three basins and allowed the process water to enter the mixed media filters at pH 7.4. Again with the extra capacity built into the system the reduction of pH from 10.5 to below 8.0 was easily controllable. The carbon dioxide used during most of the project was purchased and was not that produced by the recalcination furnace. The carbon dioxide used was 98% pure CO_2 which was diluted by the flue gas compressor to approximately 10% CO_2 concentration.

The results from the chemical clarification system are presented in Tables 19 through 22 and in Table 26. As shown in Table 21, the lime clarification system reduced the phosphorus concentration to 0.56 mg/l as P. This is an 80% capture of phosphorus that entered the clarifier system.

TABLE 26

REMOVALS OF CHEMICAL OXYGEN DEMAND (COD)
AND TOTAL ORGANIC CARBON (TOC) DURING
EVALUATION OF THE LOW LIME PROCESS

	COD mg/l	TOC mg/l as C
Raw	---	---
Primary	---	---
Secondary	32.54	16.0
Lime Clarified	25.45	11.6
Filtered	17.03	9.6
Carbon Adsorption	9.34	3.6

Following the chemical clarification system, the effluent was fed to dual media filters. Five filters were in continuous operation with one additional unit either in backwash or standby. The flow as shown in Table 23 averaged 2.69 gpm/sq ft. The filters were backwashed on a 24-hour cycle at an average headloss of 8 feet.

As seen in Table 20 the dual media filters removed suspended solids for and average capture of 5.6 mg/l per unit. As mentioned in the high lime process discussion, an accurate efficiency of the filters could not be determined because of the undetermined portion of the calcium carbonate that was solubilized during pH adjustment.

Following filtration the filtrate was adjusted to pH 7.4 by the addition of carbon dioxide in the stabilization basin and pumped to the three parallel activated carbon adsorption systems. As seen in Table 23, the hydraulic loadings of the three systems averaged 5.05, 4.39 and 4.31 gpm/sq ft, respectively. The flows through the individual units were controlled successfully as instrumentation was refined and additional equipment was added. The lead column of each system was backwashed every 24 hours with the final columns being backwashed every 48 hours. A total of 24,000 gallons of water was required to backwash a carbon column as compared to 39,000 gallons to backwash a dual media filter. During the backwash of a column, treatment by only one column of that particular train was provided.

The performances of the activated carbon system based on the composite samples of the three trains are presented in Tables 19 through 22 and in Table 26. The results are the averages for the 89 days of operation. From data in Table 26, the units removed approximately 63% of the inlet Total Organic Carbon and 44% of the Chemical Oxygen Demand. The degree of nitrification in the activated carbon columns, from 4.38 to 5.16 mg/l of NO_3-N, (Table 22) was typical of operation of the units throughout the study and was considerably lower than the nitrification produced in the high lime mode.

The performances of the individual carbon columns are presented in Tables 27, 28 and 29. The loadings to the columns, based on the cumulative flows through each column, are presented in Table 30. These loadings and cumulative flows, except for column T-14, included the high lime process evaluation. Column T-14 was regenerated during the plant shut down between the high and low lime modes.

The waste solids from the clarifier were pumped to the gravity thickener with a solids concentration generally between 1-5%. The concentration varied as a function of flow which averaged 54,000 gal/day. As discussed in the high lime process section, it was difficult to accurately sample and analyze the sludge because of errors associated with the highly concentrated waste stream. Therefore, the chemical analyses of the liquid streams were considered to be more reliable because of the accuracy of the chemical analyses in combination with good flow measurements. As shown in Table 31, the average solids captured in the clarifier was 5509 lb/day. The inerts in the calculation include inert solids from the backwash cycles, centrate, recalcined lime and furnace operation.

TABLE 27

PERFORMANCE OF CARBON ADSORBER TRAIN #1
DURING EVALUATION OF THE LOW LIME PROCESS

Average Flow 1.285 mgd

Cumulative flow to July 8, 1974 *	T-15	322.860 mil gal
	T-14	122.953 mil gal

Total Organic Carbon (TOC)	mg/l as C	% Removed
Influent	10.2	---
Intermediate	5.2	49.0
Effluent	3.7	63.7

Biochemical Oxygen Demand (BOD)	mg/l	
Influent	6.8	---
Intermediate	2.6	61.8
Effluent	2.4	64.7

Chemical Oxygen Demand (COD)	mg/l	
Influent	18.1	---
Intermediate	10.9	39.8
Effluent	7.8	56.9

*Note:
Adsorber column T-14 was regenerated during January 1974 and the sequence was switched to place adsorber T-15 in the lead. Column T-14 operates as the number 2 unit.

TABLE 28

PERFORMANCE OF CARBON ADSORBER TRAIN #2
DURING EVALUATION OF THE LOW LIME PROCESS

Average Flow 1.117 mgd

Cumulative flow to July 8, 1974*

	T-16	244.329 mil gal
	T-17	284.679 mil gal

Total Organic Carbon (TOC)	mg/l as C	% Removed
Influent	10.0	---
Intermediate	5.4	46.0
Effluent	4.2	58.0

Biochemical Oxygen Demand (BOD)	mg/l	
Influent	6.6	---
Intermediate	2.5	62.1
Effluent	2.2	66.7

Chemical Oxygen Demand (COD)	mg/l	
Influent	18.0	---
Intermediate	10.3	42.8
Effluent	10.2	43.3

* Note:
Carbon adsorber T-16 was taken out of service for a short period of time to repair a faulty valve.

TABLE 29

PERFORMANCE OF CARBON ADSORBER TRAIN #3
DURING EVALUATION OF THE LOW LIME PROCESS

Average Flow 1.095 mgd

Cumulative flow to July 8, 1974		T-18	225.981 mil gal
		T-19	225.981 mil gal
Total Organic Carbon (TOC)	mg/l as C		% Removed
Influent	10.1		---
Intermediate	5.3		47.5
Effluent	3.9		61.4
Biochemical Oxygen Demand (BOD)	mg/l		
Influent	6.6		---
Intermediate	2.4		63.6
Effluent	2.2		66.7
Chemical Oxygen Demand (COD)	mg/l		
Influent	18.3		---
Intermediate	12.4		32.2
Effluent	8.8		51.9

TABLE 30

CUMULATIVE COD AND TOC LOADINGS
AT THE END OF THE LOW LIME EVALUATION*

	lbs COD/lb Carbon	Cumulative Flow, mil gal	Lb TOC/lb Carbon
Set 1			
T-15	.19299	322.860	.08594
T-14**	.08128	122.953	.02605
Set 2			
T-16***	.22266	244.329	.11206
T-17	.12755	284.679	.05709
Set 3			
T-18	.23164	225.981	.12044
T-19	.09940	225.981	.02576

* Loadings cumulative since columns were placed in service and include the high lime evaluation except for T-14.
** Regenerated December 1973
*** T-16 was out of service due to a malfunctioning valve for approximately one and one-half months.

TABLE 31

DAILY SOLIDS PRODUCTION FOR THE LOW LIME EVALUATION

Total solids production in the clarifier based on an average flow of 4.68 mgd

$CaCO_3$	831 lbs/mil gal	x	4.68	=	3889 lb/day
$Ca_5OH(PO_4)_3$	98.58 lb/mil gal	x	4.68	=	461 lb/day
Suspended Solids	22.52 lb/mil gal	x	4.68	=	105 lb/day
$Mg(OH)_2$	16.93 lb/mil gal	x	4.68	=	79 lb/day
$Fe(OH)_3$	116.6 lb/mil gal	x	4.68	=	546 lb/day
Inerts	91.74 lb/mil gal	x	4.68	=	429 lb/day
			TOTAL		5509 lb/day

VIII. CARBON REGENERATION

With the failure of the two reactor clarifiers on November 18, 1973, carbon column T-14 was prepared for regeneration. As seen in Tables 13 and 16, the carbon in column T-14 was not exhausted, but because the grant period was ending it was felt that at least one regeneration should be done. The column to date had averaged 57% removal of the Total Organic Carbon (TOC) initially 7.5 mg/l as C, to produce an effluent with 3.2 mg/l as C. With the main portion of the plant out of service, the regeneration system was able to be started with full operator attention. This proved to be quite necessary as numerous problems were encountered with the carbon transfer system. Carbon adsorbers T-16 and T-18 were regenerated in August 1975, and these data are also presented.

As seen in Table 32, one of the difficulties encountered in the regeneration process was obtaining an accurate measurement of the quantities of carbon involved. The amount of carbon delivered by the supplier on the original delivery for column T-14 was 72,100 pounds, however, based on physical measurements and assuming a bulk density of 26 lb/cu ft, it was calculated that 73,000 pounds was delivered, a difference of 1.2%. Table 33 shows the operational data. Variations in the regeneration procedures were numerous as means of transferring and measuring the carbon were being developed. The daily amounts of carbon in the spent and regenerated carbon storage tanks were difficult to determine because of uneven carbon levels and the inaccessibility of the tanks. An accurate measurement of the losses due to regeneration could therefore not be determined with the total system and measuring options available. The closest estimate of losses that can be given from the three regenerations is that losses were between 8 and 10%.

TABLE 32

INVENTORY OF CARBON

	T-14	T-16	T-18
Carbon reported delivered by supplier, lb	72,100	70,640	69,640
Carbon as measured in column, lb*	73,000	69,430	68,888
Carbon before regeneration, lb*	69,792	69,800	68,500
Carbon fed to the furnace based on feed rate, lb	65,112	67,200	61,500

*Note, calculations based on 26 lb carbon/cu ft.

As compared in Tables 33 through 35, the regenerations of columns T-16 and T-18 resulted in lower iodine numbers in the regenerated carbon because of higher loadings for the spent carbon and increased feed rates to the regeneration furnace. The regeneration process was controlled by the results from the apparent bulk density test. This test was the only one in which the results could be received immediately and corrections made to the process. As can be seen in Table 34, the densities also reflected the increased feed rates and resulting decreased regeneration efficiency. Iodine tests along with determination of ash concentrations were performed in the laboratory and normally required 5-7 days before the results were available. These numbers, therefore, could not be used for timely adjustment of operating parameters for carbon regeneration. Table 36 shows one sieve analysis from column T-18 indicating some breakdown in particle size due to the regeneration and transfer processes. Table 37 gives the natural gas consumption in the three regenerations and the maximum temperatures through which the carbon passed during regeneration. The hottest hearth was the bottom hearth, #4.

The three carbon columns which were regenerated represent the startup and debugging phases of operation. Numerous modifications and corrections would have to be made to the delivery and measuring operations before carbon losses could be accurately determined.

TABLE 33

OPERATING DATA FOR REGENERATION OF CARBON

	T-14	T-16	T-18
Total feed to furnace, lb	65,112	67,200	61,500
Total operating days	16	15	12
Average hours of feed/day	16.3	9.3	10.8
Carbon feed rate, lb/min	4.14	8.03	7.9
Transfer rate, lb carbon/gal water	0.47	.51	1.0
Steam rate, lb/hr	140	146	135
lb steam/lb carbon	.56	.30	.28

TABLE 34

RESULTS OF LABORATORY ANALYSES OF CARBON

	T-14	T-16	T-18
Iodine number, virgin	950	950	950
Iodine number, spent	748	428	554
Iodine number, regenerated	936	741	747
Ash number, spent, %	5.0	8.2	7.3
Ash number, regenerated, %	6.1	8.9	9.3
Apparent bulk density, virgin, grams/cc	.500	.500	.500
spent, grams/cc	.552	.572	.565
regenerated, grams/cc	.486	.523	.516

TABLE 35

CARBON LOADINGS AT TIME OF REGENERATION

	T-14	T-16	T-18
TOC loading ($X10^{-2}$ $\frac{\text{lb TOC}}{\text{lb carbon}}$)	9.6	18.5	19.5
BOD loading ($X10^{-2}$ $\frac{\text{lb BOD}}{\text{lb carbon}}$)	2.4	16.5	18.9
COD loading ($X10^{-2}$ $\frac{\text{lb COD}}{\text{lb carbon}}$)	21.1	41.0	46.3

TABLE 36

SIEVE ANALYSES OF CARBON FROM T-18 CARBON ADSORBER

Sieve size/U.S. No.	Spent (% retained)	Regenerated (% retained)
4	2.60	0
8	9.99	4.25
10	17.71	24.69
12	19.49	21.77
14	19.67	17.71
16	12.96	11.28
20	11.96	11.08
30	4.37	3.07
40	.70	.40
325	.55	.75

TABLE 37

FURNACE CONDITIONS DURING CARBON REGENERATION

	T-14	T-16	T-18
Natural gas actually used, cu ft/hr	1494	1465	1512
Temperature (max.hearth)(°F)	1677	1718	1800
Temperature (afterburner) (°F)	1375	1344	1215
Ratio of natural gas to carbon cu ft natural gas/lb of carbon	6.0	3.0	3.2

IX. COST ANALYSIS

The major purpose of this project as conceived by EPA and WSSC was to gather reliable data with regard to operating costs and operational problems. Most of EPA's previous work in the Physical-Chemical AWT processes had been accomplished on a small pilot scale and the purpose of the Piscataway Model Plant project was to build and evaluate a larger system. The following pages show cost breakdowns and mention several operating problems encountered. Costs for the Model Plant were high for the following reasons:

1. Data are based on the startup period where equipment was being modified and adjusted.
2. Operators were unfamiliar with the plant. The majority of the staff had little training in the wastewater field or prior experience in treating wastewater.
3. Processes were not operated within optimum ranges of efficiency.

The caliber and quality of the operating staff should be a major consideration when discussing and selecting unit processes for the facility. The startup period at the Piscataway Model AWT Plant included numerous problems revolving around three unfortunate situations. First was the inability to hire and retain qualified personnel. Secondly, numerous mechanical problems were further complicated when a fire broke out late in the construction phase, causing extensive smoke damage to most of the pneumatic controls and automatic equipment, as well as creating delays in completion of construction. Third was the lack of redundancy in key pieces of equipment.

The basis for design was that the plant would be a model facility and have a five-year design life. This was done with the concept of obtaining as much technical data as possible while keeping the capital cost at a minimum. This, however, resulted in design deficiencies that caused operating problems that might not occur in a plant designed for a longer life and for greater operating efficiency. Prior to startup in late 1972, the new engineer and his staff of ten plant operators attended daily classroom training sessions and performed actual plant work before entering the Model Plant. This period included spending more than 40 hours at the EPA Blue Plains Pilot Plant to gain vital operating experience. The operators were also given practical experience while working at the Piscataway Secondary Treatment Plant. Although these eleven persons began the initial testing and startup in September 1972, the plant had lost the trained engineer and 7 of the original 10 operators by December 1972. During the next year a total of 51 operators and 3 engineers-in-charge had been assigned to the plant at one time or another. This high turnover combined with both normal construction and design defects resulted in a long startup period. One of the major lessons learned from the Model Plant testing was that quality

personnel are needed if a relatively complex plant is to be operated on a continuous basis.

Capital Expenditures

As mentioned previously, the concept of a Model Plant began with EPA in 1966. The original talks were based on what size plant could be built for approximately 2.5 million dollars. However, following a preliminary design study it was determined that a 5 mgd plant could be constructed for 3.2 million dollars and the project proceeded into design and construction. As shown in Table 38, the final cost of the Model Plant was 4.68 million dollars. Table 39 shows a more detailed breakdown of the costs in which the EPA's final share was 3.1 million dollars, the State of Maryland paid 0.3 million and WSSC paid 1.1 million dollars.

Tables 40 and 41 give a cost breakdown for the Piscataway Secondary Plant. Phase I was for the 5 mgd system which is used to treat the sewage entering the Model Plant. The total cost of the 30 mgd Secondary Plant was 13.5 million dollars. Table 42 gives the breakdown on engineering services supplied to EPA and WSSC by Roy F. Weston, Inc.

Tables 43 through 45 give breakdowns of the Model Plant expenses for the unit processes. Note that in Table 43 an effort has been made to estimate the costs of a 5 mgd plant if the low lime concept were to be the basis for design. Also in Table 43, note that the engineering services have been included as complete plant costs and are not broken down for the individual processes.

TABLE 38

CAPITAL COSTS OF THE MODEL PLANT

General Contractor (Main Contract)		$3,037,100
Change Orders		210,338
Furnace Contractor		529,000
Activated Carbon		148,356
Centrifuges		66,480
Electric Substation		153,800
	Sub Total	$4,145,074
Engineering Services		535,243
	Total	$4,680,317

TABLE 39

DISTRIBUTION OF CAPITAL COSTS OF THE MODEL PLANT

U.S. Environmental Protection Agency	
Under Project 17080DZY	
Research and Development	EPA's share 75%
$3,200,000 x .75	= $2,400,000
Under Project WPC - MD - 233	
Additional Facilities	$1,320.453
55% EPA's share	
$1,320,453 x .55	= $ 726,249
25% State of Md. share	
$1,320,453 x .25	= $ 330,113
Washington Suburban Sanitary Commission	
Main Contract	$ 800,000
Additional Facilities	$ 264,090
	$1,064,090

TABLE 40

WSSC CONSTRUCTION PHASES FOR THE PISCATAWAY SECONDARY TREATMENT PLANT

PHASE	COST	CONSTRUCTION BEGUN	CONSTRUCTION COMPLETED	REMARKS
I	$ 3,027,000	1966	1967	5 mgd complete secondary plant and solids handling facilities.
II	$ 2,510,435	1968	1970	25 mgd expansion of aeration basin and secondary clarifiers.
III	$ 2,431,634	1970	1974	25 mgd expansion of grit chamber, primary clarifier and thickening areas.
IV	$ 1,079,338	1970	1971	Four polishing ponds installed improved process effluent.
V	$ 1,702,500	1972	1974	Construction of additional solids handling facilities including a vacuum filter and three incinerators.
VI	$ 1,768,762	1972	1974	Construction of additional solids handling including vacuum filter and incinerator. (In conjunction with Phase III).
VII	$ 967,317	1973	1975	Modification to existing plant including samplers and misc. safety equipment.
TOTAL	$13,486,986			

TABLE 41

COST BREAKDOWN FOR THE

30 MGD SECONDARY TREATMENT PLANT

Project I			
5 mgd			
	Primary	$ 774,078	
	Secondary	$ 1,501,907	
	Solids Handling	$ 751,015	
		Total Cost	$ 3,027,000
Project II			
	Ponds	$ 775,621	
	Other	$ 302,518	
		Total Cost:	$ 1,079,339
Project III			
Expansion 30 mgd			
	Primary	$ 1,495,313	
	Secondary	$ 2,748,209	
	Solids Handling	$ 4,796,426	
		Total Cost	$ 9,039,948
			$ 13,146,287
WSSC Phases	I	$ 3,027,000	
	II	$ 2,510,435	
	III	$ 2,431,634	
	IV	$ 1,079,338	
	V	$ 1,702,500	
	VI	$ 1,768,762	
	VII	$ 967,317	
		$ 13,486,986	

Note: Difference between two totals is $340,669 which includes expenses for the Model Plant, miscellaneous landscaping, and road work.

TABLE 42

COSTS OF ENGINEERING SERVICES FOR THE MODEL PLANT

PRELIMINARY ENGINEERING DESIGN AND REPORT	$ 28,449
STUDY OF ADVANCED TREATMENT	7,281
PREPARATION OF PLANS AND SPECIFICATIONS	288,511
CONSTRUCTION SERVICES	190,095
PREPARATION OF OPERATIONS MANUAL	20,907
Total	$535,243

TABLE 43

CAPITAL COSTS FOR THE MODEL PLANT UNIT PROCESSES

Process	High Lime Cost		Low Lime Cost*
Lime Clarification	$1,340,190		$ 991,000
Filtration & Carbon Adsorption	1,150,354		1,150,354
Solids Handling	1,099,247		291,930
Carbon Regeneration	401,483		401,483
Sub Total	$3,991,274		$2,834,767
Electrical Substation	153,800	**	109,198
Engineering Services	535,243	**	380,022
Total	$4,680,317		$3,323,987

*Low lime cost is based on a calculated estimate.
**Both of these numbers are proportional to the sub totals.

TABLE 44

BREAKDOWN ON CAPITAL COSTS FOR THE MODEL PLANT UNIT PROCESSES

Clarification		
	Excavation	$ 52,700
	Concrete Work	187,400
	Mechanical Work	237,000
	Electrical Work	28,000
	Painting	10,000
	Other Work	20,300
		$ 535,400
Filtration		
	Steel Shelves	$ 118,000
	Underdrains	35,000
	Filter Media	8,000
	Mechanical Work	29,500
	Piping and Valves	32,000
	Painting	3,000
	Electrical Work	600
	Other	7,100
		$ 232,200
Carbon Adsorbers		
	Steel Tanks and Supports	$ 232,000
	Filter Bottom	28,000
	Mechanical Work	26,000
	Piping and Valves	75,000
	Electrical Work	6,000
	Painting	7,900
	Other Work	1,800
		$ 367,700
Operations Building		$ 1,076,900
Centrifuge Area & Cake Handling System (excluding centrifuge)		33,800
Lime Handling System		100,200
$FeCl_3$ and Polymer System		43,200
Motor Control Center and Power Distribution		105,500
Instrumentation		308,800
Miscellaneous Equipment, Steel and Clean-up		226,337
		$ 3,031,037

TABLE 45

CAPITAL COST BREAKDOWN FOR THE MODEL PLANT EQUIPMENT

GENERAL CONTRACT	COST	DATE OF COMPLETION
Reactor Clarifier	$ 294,000	October 72
Recarbonation Basin & Associated Tanks	92,400	October 72
Lime Pumping Stations	81,000	October 72
Thickener & Pumping Station	68,000	October 72
Operations Building	1,076,900	April 72
Dual Media Filters	194,000	April 72
Backwash Tanks	78,400	April 72
Stabilization Tank	50,500	April 72
Activated Carbon Adsorbers (excluding carbon)	247,500	April 72
Carbon Storage Tanks	30,500	April 72
Lime Bins	31,500	May 72
Lime Slaker System	25,700	February 73
Assorted Lime Handling Systems	43,000	February 73
Lime Cake Handling Systems	33,800	October 72
Miscellaneous Plant Equipment	61,600	April 72
Chemical Feed System (Ferric & Polymer)	43,200	October 72
Steam Generator	13,500	April 72
Platform & Structures	69,200	April 72
Motor Control Center & Power Distribution	105,500	March 72
Instrumentation	308,800	February 73
Sitework & Piping	88,100	April 72
Sub Total	$ 3,037,100	
Change Orders	210,338	February 73
Furnace Contracts	529,000	March 73
Activated Carbon	148,356	October 72
Centrifuges	66,480	July 72
Total	$ 3,991,274	

Operational Costs

The following pages provide operating costs based on pay schedules shown in Table 46 and cost of chemicals and energy shown in Table 47. Several factors to consider when analyzing these data are:

1. The miscellaneous group used in the tables and distribution diagrams includes items such as water usage, hoses, shovels, office and lab supplies, and cleaning supplies.

2. All chemical costs are FOB at the plant.

3. Costs do not include those for laboratory personnel and for the analytical tests.

4. No overhead costs are included in the cost breakdowns other than those for the staff on hand.

As shown in Table 48, operating costs for tertiary treatment using the Low Lime Process with wasting of wet solids were 28.92¢/1000 gallons ($289/mil gal). Table 49 shows an operating cost of 32.97¢/1000 gallons for the Low Lime Process with solids dried and wasted and Table 50 shows an operating cost of 35.80¢/1000 gallons for the High Lime Process. The costs reflect the plant operation, but do not consider the cost to landfill the waste from the three types of lime treatment. With the three types of processes, the personnel requirements are the same. In certain instances, and depending on unit processes, personnel requirements can be cut for the Low Lime Processes. At the Piscataway Model Plant, the staff operated separately from the staff at the Piscataway Secondary Plant. This separation of the staffs caused increased costs. Table 51 gives the personnel breakdown in the various unit processes.

Figures 15 through 17 are breakdowns of the operating costs as presented in Tables 48 through 50. Figures 18 through 21 give distribution diagrams of the capital costs and power costs of the Model Plant. As mentioned previously, the calculations for the Low Lime Process in both the capital cost and power requirements are the best estimates that can be made from the available data.

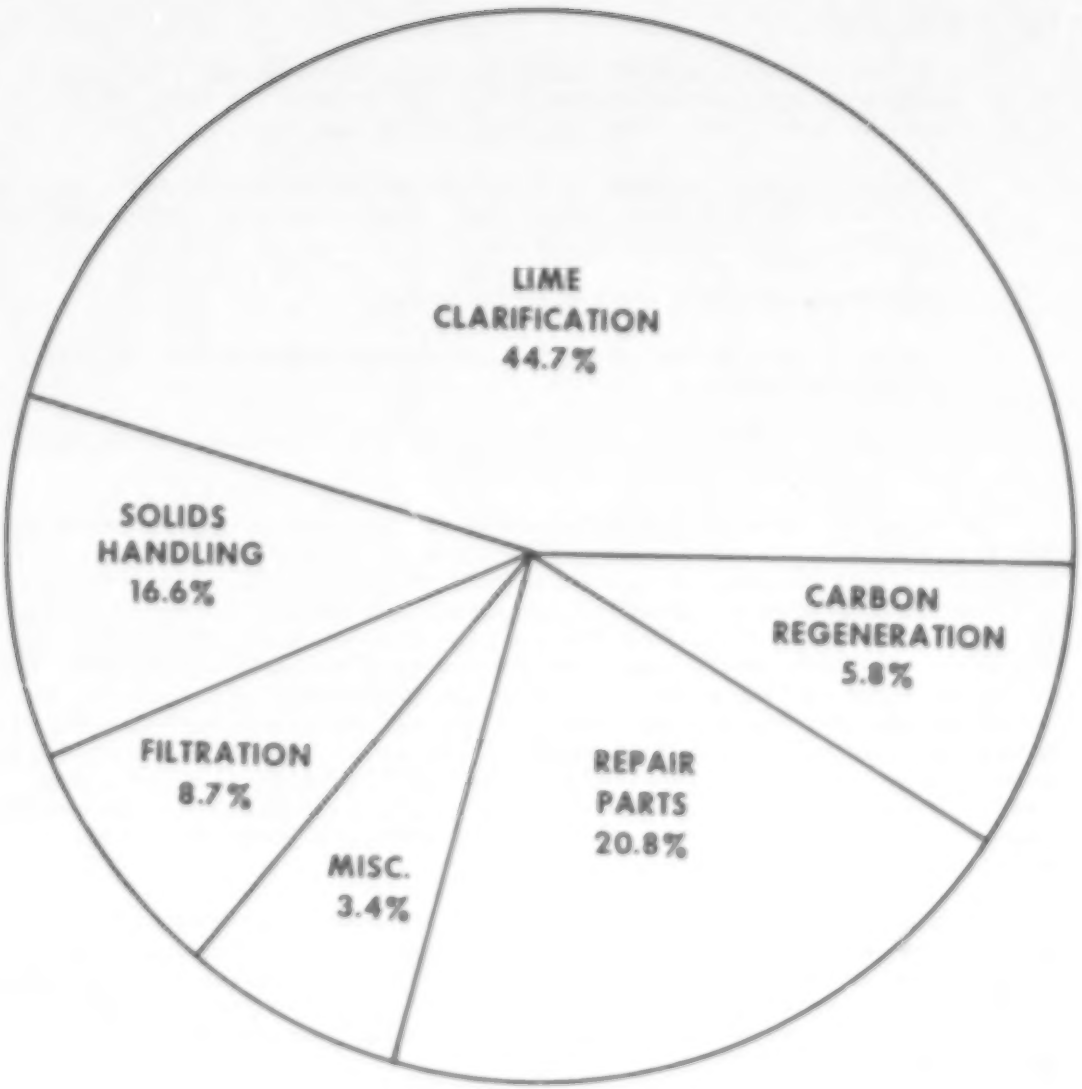

Figure 15. Distribution of operating costs for the low lime process with wasting of wet solids.

Figure 16. Distribution of operating costs for the low lime process with solids dried and wasted.

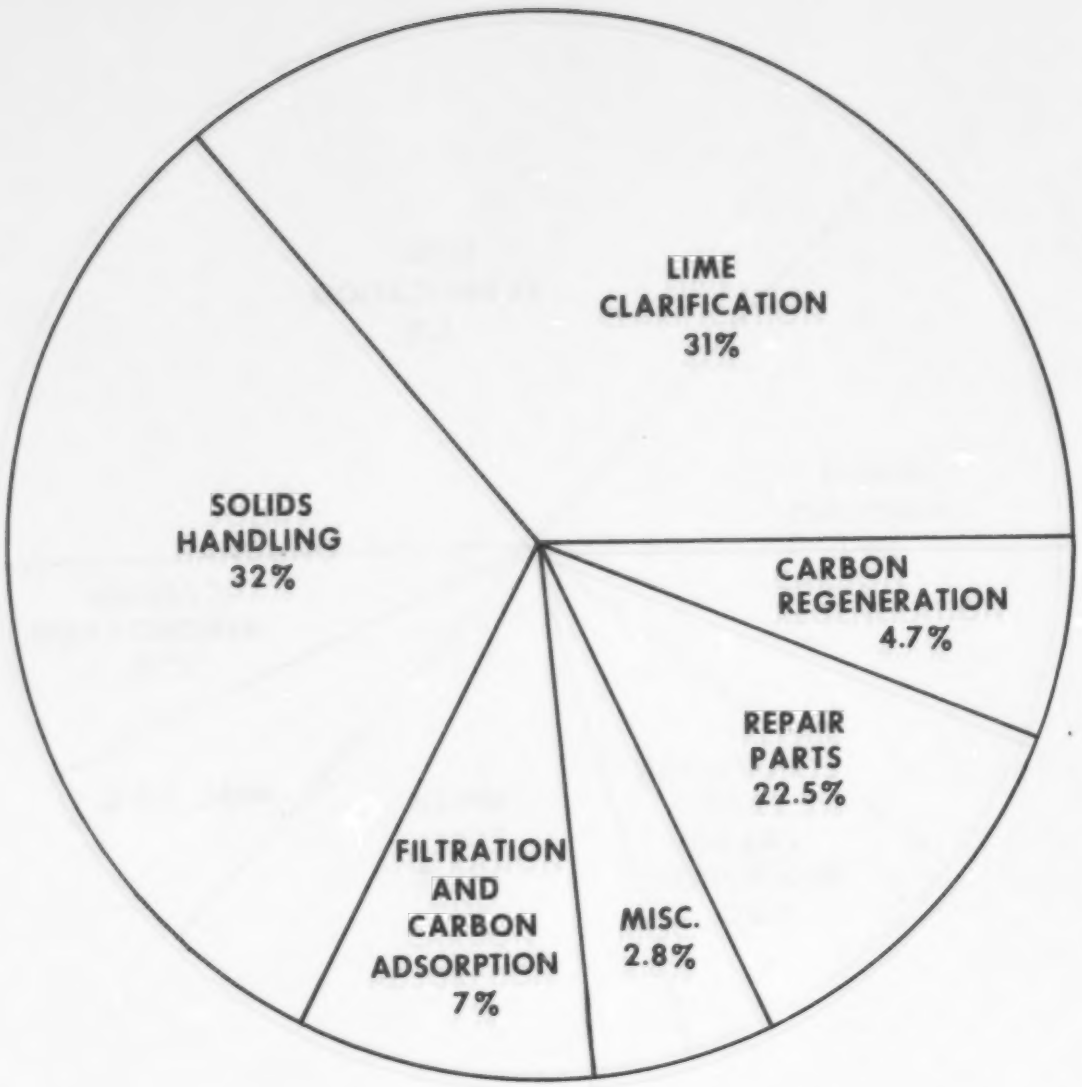

Figure 17. Distribution of operating costs for the high lime process.

Figure 18. Distribution of capital costs for the low lime process.

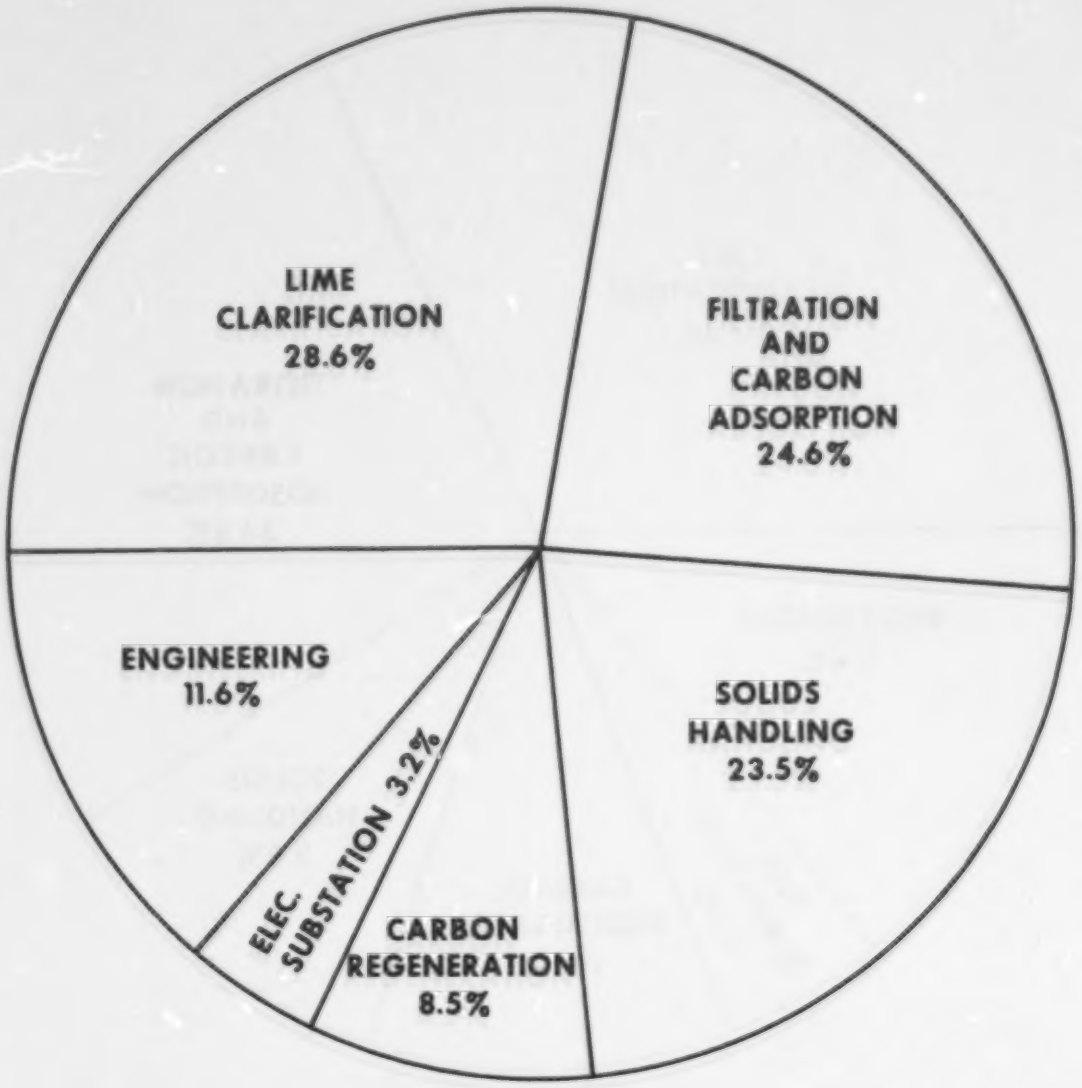

Figure 19. Distribution of capital costs for the high lime process.

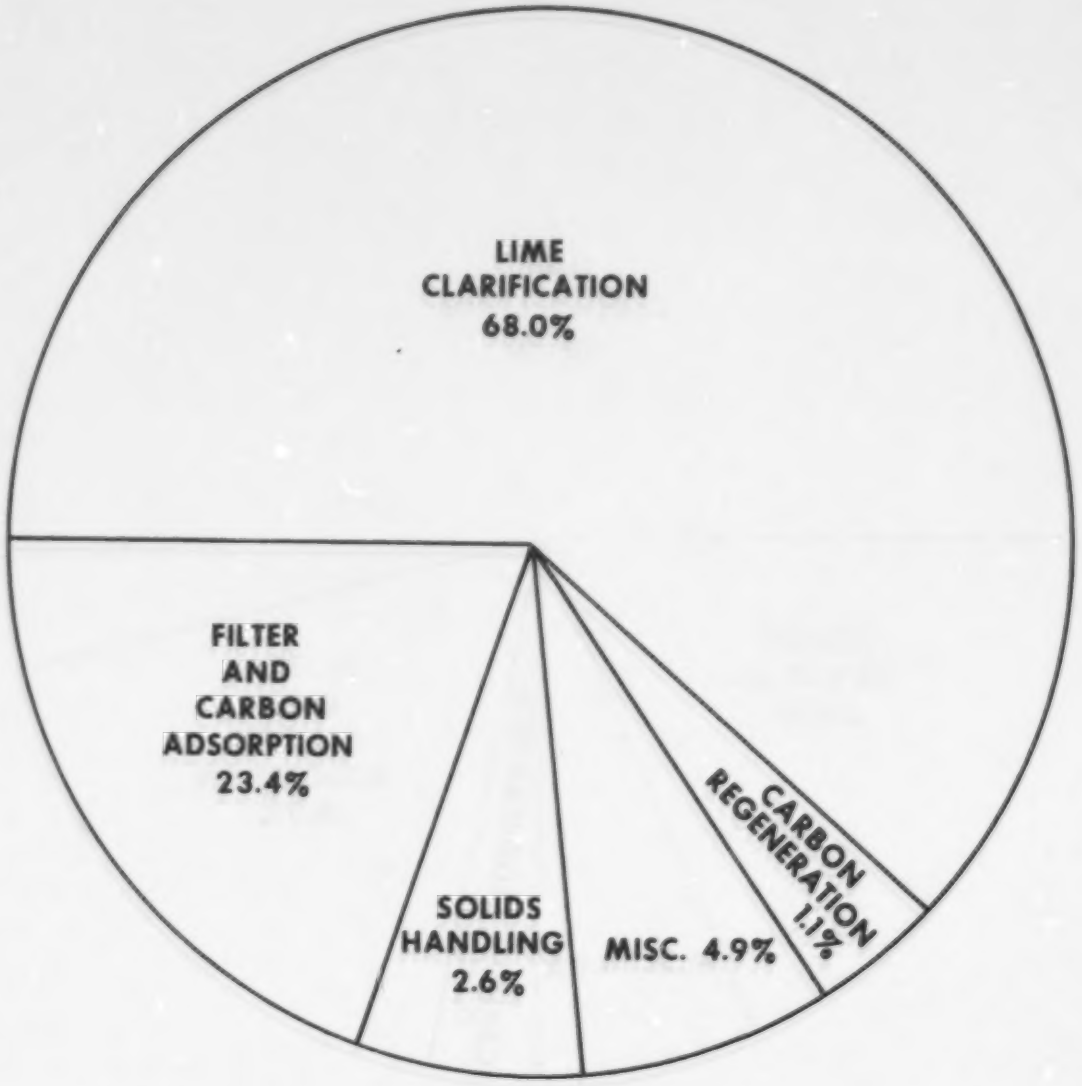

Figure 20. Distribution of power requirements for the low lime process.

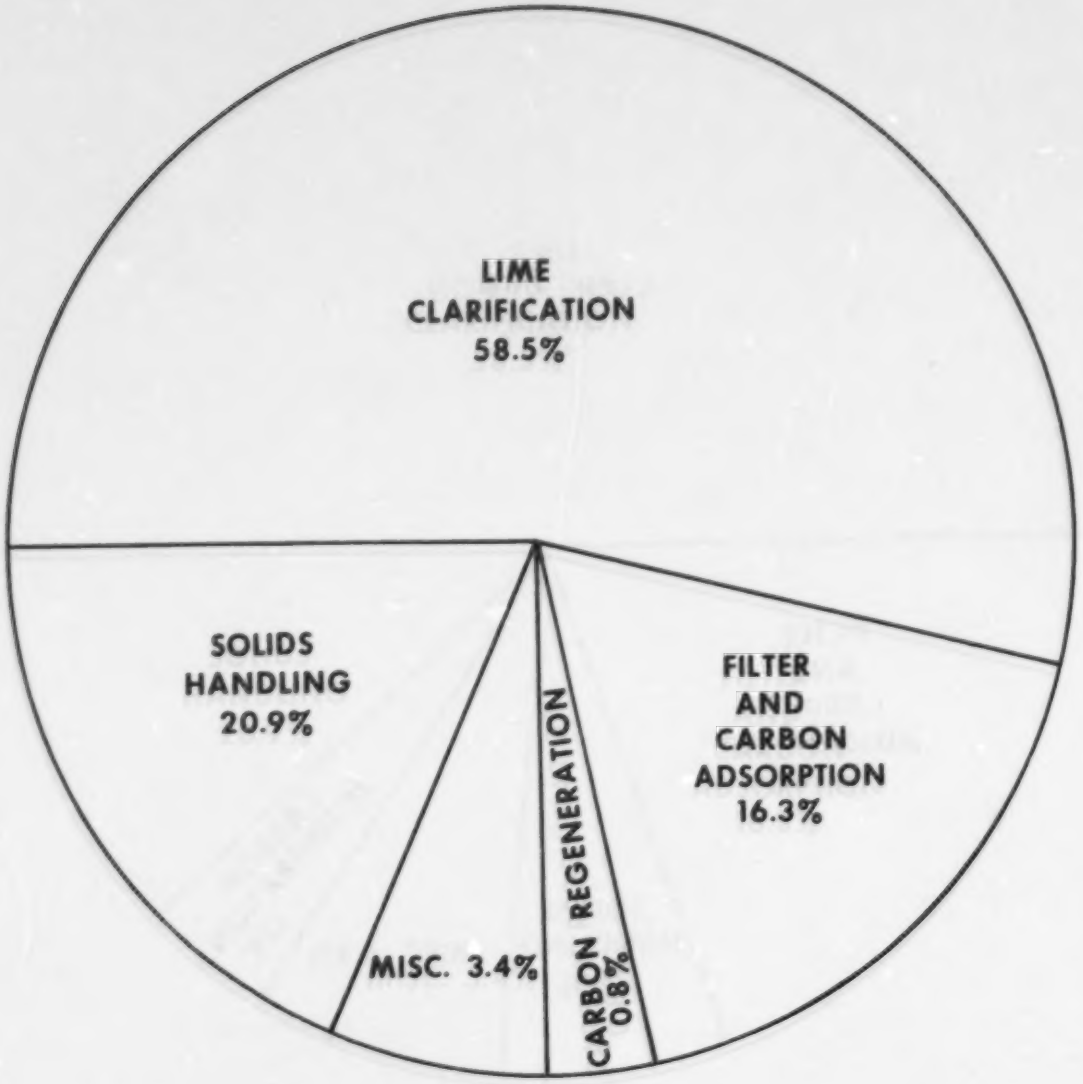

Figure 21. Distribution of power requirements for the high lime process.

TABLE 46

PAY SCHEDULES AT PISCATAWAY
Effective July 1, 1974

	Yearly Pay
Engineer II	$14,211 - $20,956
Engineer I	$12,316 - $17,317
Senior Wastewater Plant Operator	$11,710 - $14,946
Wastewater Plant Operator	$ 9,050 - $14,230
Laborer/Driver	$ 8,507 - $10,630
Chemist II	$15,410 - $19,195
Chemist I	$11,710 - $16,491
Laboratory Technician	$ 8,561 - $14,230
Bacteriologist I	$11,710 - $16,491
Electrical/Mechanical Engineers	$12,920 - $24,691
Asst. Electrical Maintenance Supervisor	$15,419 - $19,195
Electrical Mechanic	$10,379 - $13,853
Electrical Mechanic Apprentice	$ 9,547 - $11,560

TABLE 47

COST FIGURES FOR ENERGY AND CHEMICALS

Chemical	
Pebble Lime	$ 22.15/ton
Ferric Chloride	$128.50/ton
Polymer	$ 1.35/lb
Carbon Dioxide	$ 60.00/ton
Power	
Electricity	$0.0185/kWh
#2 Fuel Oil	$0.2545/gal
Natural Gas	$2.50/first 5 therms*
	$0.175/second 10 therms
	$0.165/next 15 therms
	$0.148/next 500 therms
Miscellaneous	
Water	$0.48/1000 gal
Granular Carbon	$826.61/100 cu ft
CO_2 Tank Rental	$265/month

* 1 therm = 100 cu ft

TABLE 48
Operating Costs for the
Low Lime Process With
Wasting of Wet Solids

	Cost/1000 Gallons
Lime Clarification	12.92¢
Solids Handling	4.80¢
Filtration & Carbon Adsorption	2.52¢
Miscellaneous (Water, supplies)	1.00¢
Repair Parts	6.00¢
Carbon Regeneration	1.68¢
	28.92¢/1000 gallons

TABLE 49
Operating Costs for the
Low Lime Process With
Solids Dried & Wasted

	Cost/1000 Gallons
Lime Clarification	10.22¢
Solids Handling	9.55¢
Filtration & Carbon Adsorption	2.52¢
Miscellaneous (Water, supplies)	1.00¢
Repair Parts	8.00¢
Carbon Regeneration	1.68¢
	32.97¢/1000 gallons

TABLE 50
Operating Costs for the
High Lime Process

	Cost/1000 Gallons
Lime Clarification	11.08¢
Solids Handling	11.52¢
Filtration & Carbon Adsorption	2.52¢
Miscellaneous (Water, supplies)	1.00¢
Repair Parts	8.00¢
Carbon Regeneration	1.68¢
	35.80¢/1000 gallons

TABLE 51
Personnel Breakdown
by Unit Processes

	Operators hr/day	Engineer hr/day	Laborers hr/day
Lime Clarification	27	2	6
Solids Handling	34	3	8
Filtration & Carbon Adsorption	6	1	0
Carbon Regeneration	12	2	2
Total	79	8	16

1 Engineer

15 Plant Operators

2 Laborers

18 Total Staff

APPENDIX

CALCULATION OF RECALCINATION FURNACE HEAT BALANCE

Feed rate of wet sludge	3315 lb/hr
Moisture content by weight	61.5%
Total solids by weight	38.5%
$CaCO_3$ concentration in dry solids	72.0%
% volatile in dry solids	8.0%
$Mg(OH)_2$ concentration in dry solids	12.0%
$Ca_5OH(PO_4)_3$ concentration in dry solids	9.0%

Given: Average temperature based on 41 days operation

1. Exhaust gas temperature	910°F
2. Calcined product temperature	1000°F
3. Furnace shell temperature	175°F
4. Feed inlet temperature	60°F
5. Center shaft cooling air temperature	235°F

Heat Balance Calculations

Equations			
lb/hr	H_2O	3315 lb/hr x 61.5% =	2038.7
lb/hr	Solids	3315 lb/hr x 38.5% =	1276.3
lb/hr	$CaCO_3$	1276.3 lb/hr x 72% =	918.9
*lb/hr	CaO	918.9 lb/hr x 56/100=	514.6
*lb/hr	CO_2	918.9 lb/hr x 44/100=	404.3
lb/hr	Volatile	1276.3 lb/hr x 8% =	102.1
**lb/hr	Inert Ash	1276.3 lb/hr x 20% =	255.3

* Note: Chemical Equation

$CaCO_3 \rightarrow CaO + CO_2$

100 = 56 + 44 molecular weight

** Note: Inert Ash is made up of Mg, Fe, and P Compounds.

Summary of Feed Composition

Constants

1. m = mass quantity in lb/hr
2. cp = heat capacity (specific heat)
3. H = latent heat of vaporization

I. Heat Demand - material plus reaction requiring heat input

A. Heat required to heat up H_2O, evaporate H_2O, superheat water vapor to 910°F

1. Heat up liquid water to 212°F

H_2O liquid 60°F → H_2O liquid 212°F

$Q = m C_p \Delta T$

$= 2038.7 \text{ lb/hr} \times 1.0 \frac{\text{BTU}}{\text{lb°F}} \times (212°F - 60°F)$

$= 309,882 \text{ BTU/hr}$

2. Evaporate water at 212°F

H_2O liquid → H_2O steam

$Q = m H$

$= 2038.7 \text{ lb/hr} \times 970 \frac{\text{BTU}}{\text{lb of } H_2O}$

$= 1,977,539 \frac{\text{BTU}}{\text{hr}}$

3. Superheat steam to exhaust gas temperature

H_2O steam 212°F → H_2O steam 1000°F

$Q = M \ c_p \ \Delta T$

$= 2038.7 \text{ lb/hr} \times .5 \frac{\text{BTU}}{\text{lb°F}} \times (1000°F - 212°F)$

$= 803,248 \text{ BTU/hr}$

Total heat requirement of water

309,882
1,997,539
803,248
3,090,669 $\frac{\text{BTU}}{\text{hr}}$

B. Heat required to break $CaCO_3$ solid → CaO solid + CO_2 gas

$Q = m H$

$= 918.9 \text{ lb/hr} \times 785 \frac{\text{BTU}}{\text{lb } CaCO_3}$

$= 721,336 \frac{\text{BTU}}{\text{hr}}$

C. Heat required to heat up product CO_2 to $910^{\circ}F$

$$Q = m_2 \; c_p \; \Delta T$$
$$= 404.3 \text{ lb/hr} \times .25 \text{ BTU/lb}^{\circ}\text{F} \times (910^{\circ} - 60^{\circ}\text{F})$$
$$= 85{,}914 \frac{\text{BTU}}{\text{hr}}$$

D. Heat required to heat up CaO and ash to calciner outlet temperature

$$Q = m_3 \; c_p \; \Delta T$$
$$= (514.6 + 255.3) \text{ lb/hr} \times .3 \frac{\text{BTU}}{\text{lb}^{\circ}\text{F}} \times (1000^{\circ}\text{F} - 60^{\circ}\text{F})$$
$$= 769.9 \text{ lb/hr} \times .3 \frac{\text{BTU}}{\text{lb}^{\circ}\text{F}}$$
$$= 217{,}112 \text{ BTU/hr}$$

E. Radiation loss

1. Furnace Shell — Diameter 18.75 feet; Height 22.00 feet

Total surface area of furnace = 1852 sq ft

Average heat loss at $175^{\circ}F$ skin temperature is $250 \frac{\text{BTU}}{\text{hr sq ft}}$

$$Q = 250 \frac{\text{BTU} \times 1852 \text{ sq ft}}{\text{hr sq ft}}$$
$$= 463{,}000 \frac{\text{BTU}}{\text{hr}}$$

2. From center shaft cooling air

Air $_{60^{\circ}F} \rightarrow$ Air $_{235^{\circ}F}$

$$Q = m_4^{*} \; c_p \; \Delta T$$
$$= 3500 \text{ cu ft/min.} \times 60 \text{ min/hr} \times .02 \frac{\text{BTU}}{\text{cu ft}} \times (235^{\circ}\text{F} - 60^{\circ}\text{F})$$
$$= 735{,}000 \frac{\text{BTU}}{\text{hr}}$$

* Note: air volume through center shaft

Total radiation loss - 463,000 + 735,000 = 1,198,000 $\frac{\text{BTU}}{\text{hr}}$

Total required heat input

Heat (H_2O)	3,090,669
Endothermic	721,336
Product CO_2	85,914
Lime and Ash	217,112
Radiation	1,198,000
	5,313,031 $\frac{\text{BTU}}{\text{hr}}$

II. Heat Release

1. Volatile release 102.1 lb/hr x 10,000 $\frac{BTU}{lb}$ = 1,021,000 $\frac{BTU}{hr}$

2. Heat required from fuel oil

Total heat required	5,313,031 $\frac{BTU}{hr}$
Heat release volatile material	1,021,000 $\frac{BTU}{hr}$
	4,292,031 $\frac{BTU}{hr}$

#2 fuel oil thermal output = 138,000 BTU/gal*

assuming 100% efficiency

* Reference: North American Combustion Handbook

Assuming 20% excess air and a flue gas outlet temperature of 910°F, the percent of gross fuel input which is available is 68%.

Therefore, total gallons of #2 fuel oil required to produce 4,292,031 $\frac{BTU}{hr}$ is equal to: $\frac{4,292,031 \frac{BTU}{hr}}{138,000 \frac{BTU}{gal} \times 0.68}$ = 45.74 $\frac{gallons}{hour}$

45.74 $\frac{gal}{hr}$ x 24 hr/day = 1098 gal/day

Actual fuel oil usage was 1169 gal/day

Therefore, the efficiency of operation is:

$\frac{1098}{1169} \times 100$ = 94%